Effect of Particle and Bed Diameter on Characteristic Velocities in Compartmented Fluidized Bed Gasifier

V.S. Chok
Department of Chemical Engineering,
Universiti Teknologi PETRONAS,
31750 Tronoh, Perak, Malaysia
chokvuisoon@yahoo.com.sg

S.K. Wee, A. Gorin, H.B. Chua
School of Engineering and Science,
Curtin University of Technology Sarawak Campus,
CDT 250, 98009 Miri, Sarawak, Malaysia
alexander.g@curtin.edu.my

Abstract—**A 500 kg/day biomass (palm shell) fed compartmented fluidized bed gasifier (CFBG) pilot plant for the purpose of synthesis gas production and power generation has been constructed in Curtin University of Technology, Sarawak Campus. The reactor ID is 66cm with 60:40 cross sectional area ratio for combustor and gasifier respectively. Each compartment consists of a pair of devices at the partitioning wall for internal solid circulation. The minimum and complete fluidization velocity (U_{mf} and U_{cf}) experiments were first conducted using pure sand only as the bed material. This is important in order to check whether the fluidization behavior, particularly the bed pressure drop profile in CFBG is distinctive from those usually observed in the fluidized bed of cylindrical shape using common bed material. Besides that, the characteristic velocities obtained provide the operating parameters for the CFBG when the presence of other bed material is very small. Despite of the unique reactor feature, the bed pressure drop profiles are closely resemble to those observed in the laboratory scale cylindrical column. Proper fluidization is attainable in both compartments for all the sand sizes used. In the present work, $U_{cf} > U_{mf}$, in both compartment and this is observed in all the sand sizes used. The U_{mf} and U_{cf} values are affected by the different particle size in the sand and bed geometry. The former leads to $U_{cf} > U_{mf}$ and U_{cf}/U_{mf} ratio is approximately constant (except for smallest sand size) for gasifier and combustor respectively, while the latter leads to $(U_{cf}, U_{mf})_{gasifier} > (U_{cf}, U_{mf})_{combustor}$ due to cumulative effects of the bed diameter and particle-to-bed diameter ratio. Based on these studies, the governing parameters can be minimized when utilizing larger sand size as bed material, hence avoiding physical modification on the vessel.**

Keywords-component; compartmented fluidized bed; minimum fluidization velocity; complete fluidization velocity

I. INTRODUCTION

Today, the world is in urgent need of alternative energy to reduce the dependency on fossil fuel and to overcome climate change. Bioenergy is receiving mass attention due to its renewable and abundance supply, climate friendly and sustainable. Biomass gasification is one of the promising technologies to produce bioenergy. In the area of biomass gasification, fluidized bed reactor is identified as the best means to handle diverse feedstock due to its effectiveness of heat and mass transfer between the gas and solid phases [1].

Presently, most of the fluidized bed gasifiers require the used of pure oxygen instead of air to generate heat for gasification reaction. This is to eliminate nitrogen contamination/dilution in the product gases. On the other hand, almost all of these technologies employ fast circulating fluidized bed reactor system (resembling to fluid catalytic cracker) that necessitates the excessive utilization of gases and/or steam. This increases further the total energy demand as additional energy is needed to elevate the temperature of the fluidizing agents to the desired condition [2].

Compartmented Fluidized Bed Gasifier (CFBG) is an internally circulating bubbling fluidized bed reactor system currently under intensive development following the construction of 500 kg/day biomass pilot plant. CFBG is a compact reactor system that consists of two compartments, i.e. combustor and gasifier, each with a pair of internal device for internal solid circulation. Air is used in the combustor to generate heat for the steam gasification reaction. The heated bed materials from the combustor acts as heat carrier is circulated internally to the gasifier and vice versa while the combustion and gasification product gases from the respective compartments are strictly separated. The bubbling fluidization in CFBG would mean a much lower utilization of fluidizing air and steam in the compartments, hence requiring significantly less energy demand. These coupling benefits of indirectly heated and bubbling fluidization make CFBG a viable alternative for biomass gasification.

CFBG has distinctive geometrical features due to the compartmentalization and the presence of internal solid circulating devices. It is therefore necessary to check whether the fluidization behavior, particularly the bed pressure drop profile in CFBG is distinctive from those usually observed in the fluidized bed of cylindrical shape using common bed material only, i.e. sand. If comparable, it then permits the used of this profile to determine the characteristic velocities i.e. the minimum and complete fluidization velocity (U_{mf} and U_{cf}). Earlier works emphasized on the fluidization quality, a ratio of experimental to theoretical bed pressure drop in CFBG using various bed materials. Interested readers are encouraged to refer [3]-[4].

In addition, it follows that the characteristic velocities obtained in this condition provide the operating parameters for the CFBG when the presence of other bed material (e.g. biomass, char etc.) is very small or negligible. Meanwhile, the

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effect of bed diameter, particle diameter and particle-to-bed diameter ratio on the characteristic velocities are also presented. Besides that, comparison between experimental characteristic velocities with correlations prediction is also included.

II. EXPERIMENTAL

A. Apparatus

A schematic of the experimental setup is illustrated in Fig. 1. The cold flow model as shown in Fig. 2 has a 0.66 ID and is divided into 2 compartments i.e. combustor and gasifier by a vertical wall in 2:1 cross-sectional area ratio [4]. A pair of devices known as v-valve and riser is available in each compartment for internal solid circulation between the respective beds. In the present studies, only the respective beds were subjected to air for fluidization. The flow rates were regulated using rotameters (measure up to 2600 liter per minute) to maintain the bubbling mode of fluidization. Pressure drops were measured using water manometers at 3 different locations for total pressure drop (ΔP_t), across distributor (ΔP_d) and bed (ΔP_b) respectively. It has been confirmed that the bed pressure drop (used to determine the characteristic velocities) is equal to the difference of total and across distributor pressure drop in all experiments.

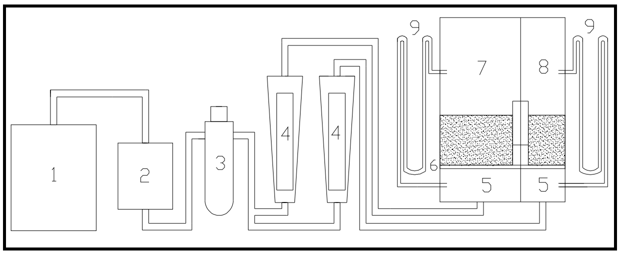

Fig. 1. Experimental setup - 1: compressor; 2: dryer; 3: pressure regulator; 4: rotameter; 5: plenum; 6: perforated distributor; 7: combustor; 8: gasifier; 9: manometer [4].

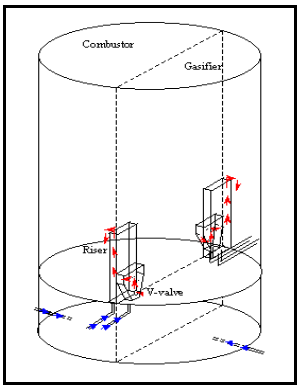

Fig. 2. Isometric view of CFBG [4].

The effective diameters, D_e computed based on (1) were found to be 25.7 and 41.3 cm for gasifier and combustor respectively. The presence of v-valve and riser in both compartments has been addressed when considering the effective bed diameter.

$$D_e = 4 \times \frac{\text{mean cross sectional area of flow channels through bed}}{\text{mean wetted perimeter of flow channels}} \quad (1)$$

B. Material

In considering the typical bed aspect ratio of 1-2, the experiments were carried out in both of the compartments at 0.4 m static bed height[1]. Large amount of bed material is used, i.e. at 77 and 101 kg respectively. 4 different types of sand sizes with density of 2,700 kg/m3 are selected as the inert materials from a nearby quarry. Table I shows the particle size distribution and mean particle diameter. Mean particle diameter, d_p for sand is computed based on

$$d_p = 1/\Sigma(x_i/d_{pi}) \quad (2)$$

where x_i is the weight fraction in the size interval, d_{pi} based on the screen analysis.

TABLE I. PARTICLE SIZE DISTRIBUTION AND MEAN DIAMETER

Sieved size (μm)	Size interval, d_{pi} (μm)	Mean particle diameter, d_p (μm)			
		196	272	341	395
		Weight fraction, x_i			
425-600	512.5	0.0023	0.0010	0.0746	0.5105
300-425	362.5	0.0085	0.5373	0.8084	0.4027
212-300	256.0	0.5924	0.3174	0.0960	0.0666
150-212	181.0	0.3236	0.1241	0.0129	0.0132
0-150	75.0	0.0732	0.0202	0.0081	0.0071

C. Procedure

The two commonly used methods for multi-component system in determining the U_{mf} and U_{cf}, i.e. via fast and slow defluidization are adopted for sand of distributed sizes [5]-[6]. Both methods are based on the bed pressure drop profiles and differing only in terms of rate of defluidization.

On U_{mf} determination using fast defluidization method, the mixture is initially fluidized vigorously ($>U_{mf}$) to maximize particles mixing and to ensure constant bed pressure drop is established, in order to form the constant fluidized bed line. Thereafter, the bed is defluidized rapidly, at bed pressure drop values below fluidized state ($<U_{mf}$), such that the mixtures uniformity remains unchanged (since particles rearrangements are avoided). This is used to form the fixed bed line. The U_{mf} is then determined from the intersection point between the fixed bed and constant fluidized bed lines. The underlying reason in fast defluidization procedure is to maximize bed homogeneity when determining the U_{mf} value. This is to establish "a wholly mixed bed" condition analogically to being a "monodisperse bed" [5].

In slow defluidization, a method used to determine U_{cf}, the approach is to allow gradual changes from fluidized bed condition to fixed bed state. U_{cf} is determined from the point when the bed pressure drop is constant [5]-[6]. The latter method is not subjected to initial fixed bed arrangement and defluidization rate [7].

Indeed, only when the operating superficial velocity greater than U_{cf} value is employed in this work to ensure all the mixture components are completely fluidized. Nonetheless, both the U_{mf} and U_{cf} are closely related to the

[1] Experiments using static bed height of 0.3 – 0.5m performed in both compartments confirmed the same results.

mixing/segregation state of wide size distribution powder as described here. In addition, these also allow comparative studies to be carried out on the various published correlations.

III. Results And Discussions

The typical bed pressure drop profile is shown in Fig. 3, obtained from quartz sand of 395µm in the gasifier. It demonstrates that:

(1) the bed pressure drop reaches identical steady state value at fluidized condition in both methods

(2) the existence of the U_{mf} and U_{cf} values

Similar bed pressure drop profile is obtained in the combustor and for all other sand sizes in both compartments. Hence, it is confirmed that the bed pressure drop profile obtained from CFBG is very similar to those from a cylindrical column of laboratory scale. Beside, the methodology used to determine the U_{mf} and U_{cf} values can be implemented for the present system despite of the unique geometrical features of CFBG.

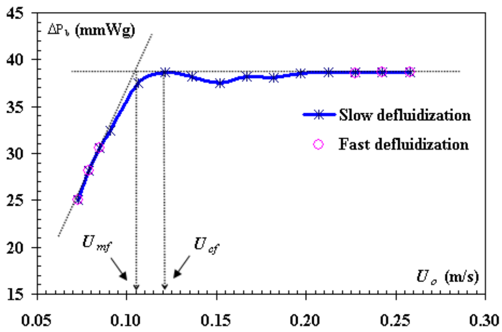

Figure 3. Typical sand bed pressure drop profile for U_{mf} and U_{cf}

A. *Effect of Particle Diameter*

The U_{mf} and U_{cf} values for both compartments with various sand sizes are presented in Table II. It can be seen that both U_{mf} and U_{cf} values increases with the increases in sand sizes. In the present work, $U_{cf} > U_{mf}$, in both compartment and this is observed in all the sand sizes used. This is due to the different particle sizes in the sand (refer to Table I for sand size distribution in the present system). Similar trends were also reported in the extensive studies by M. Pilar Aznar et al. [8], obtained from columns with diameters of 14 and 30cm.

TABLE II. U_{mf} AND U_{cf} VALUES WITH VARIOUS SAND SIZES

Compartment	Gasifier		Combustor	
Mean particle size, d_p(µm)	U_{mf} (m/s)	U_{cf} (m/s)	U_{mf} (m/s)	U_{cf} (m/s)
196	0.032	0.042	0.021	0.030
272	0.077	0.087	0.053	0.064
341	0.102	0.117	0.073	0.085
395	0.105	0.121	0.083	0.100

According to Gauthier et al. [9], who conducted the experiments in a cylindrical column of laboratory scale, whether in narrow range or Gaussian distribution, $U_{cf}/U_{mf} = 1$. In fact, $U_{cf} > U_{mf}$ for other types of distribution. Consequently, the general expectation where $U_{mf} = U_{cf}$ for single component system, a condition sometime used to validate binary mixtures correlations at the minimum end (where the second component does not present or $x_2 = 0$ wt%) is questionable; instead it should be conditional, i.e. depending on the particle size distribution [10].

Generally, the U_{cf}/U_{mf} ratio provides an indication on the mixing/segregation state of the bed material [5]. As shown in Fig. 4, except for the finest sand (196µm), the U_{cf}/U_{mf} ratio is approximately 1.15 for the gasifier and combustor. Hence, the U_{cf}/U_{mf} ratio is nearly independent of the mean particle diameter, a conclusion that was also reported in [9]. On the other hand, it can be said that the three larger particles have better mixing characteristic as compared to finest sand (with $U_{cf}/U_{mf} = 1.4$). It is known that in wide size distribution powder, the present of finer particle with those larger ones in greater portion improves the bed fluidity. In some cases, the addition of fine particle is necessary to obtain smooth bed operation [11].

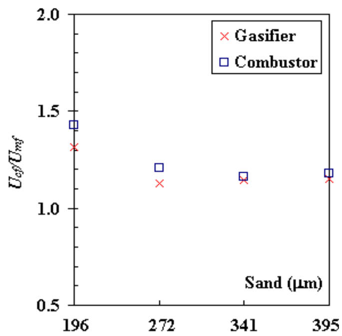

Figure 4. U_{cf}/U_{mf} ratio for various sand sizes.

B. *Effect of Bed Diameter*

Fig. 5 shows that both the U_{mf} and U_{cf} values in both compartments at various sand size. It can be seen that both the U_{mf} and U_{cf} values in the gasifier are always larger than those in the combustor. Gasifier is smaller than combustor of about 61% in effective diameter and 78% in cross-sectional area. Some specific studies in [12] on the effect of bed diameter on U_{mf} have demonstrated that the characteristic velocity increases with the decrease of bed diameter.

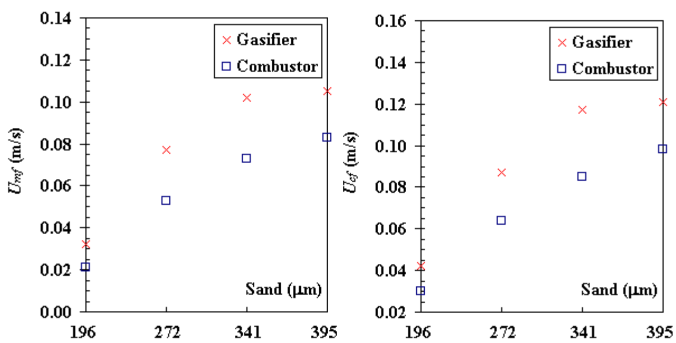

Figure 5. U_{mf} and U_{cf} values for various sand sizes in gasifier and combustor

C. Effect of Particle-to-Bed Diameter Ratio

Table III shows that these characteristic velocities ratio of gasifier-to-combustor are between 20 – 50%.

TABLE III. CHARACTERISTIC VELOCITIES RATIO

Mean particle size, d_p (µm)	$\frac{U_{mf}^{Gasifier}}{U_{mf}^{Combustor}}$	$\frac{U_{cf}^{Gasifier}}{U_{cf}^{Combustor}}$
196	1.52	1.40
272	1.45	1.36
341	1.40	1.38
395	1.27	1.21

However, these differences reduce with the increase in the mean sand diameter as shown in Fig. 6.

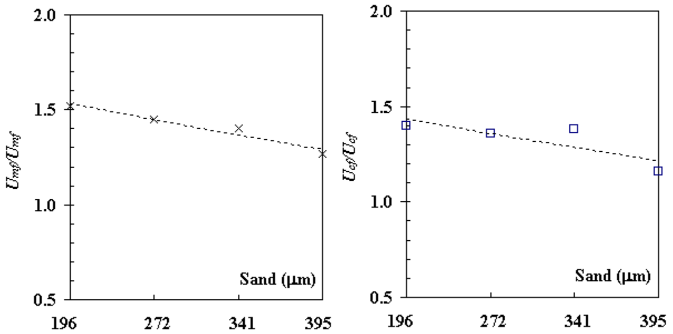

Figure 6. Characteristic velocities ratio of gasifier-to-combustor

The particle-to-bed-diameter relationship can be described by taking the dimensionless ratio of the Reynolds number for the particle and column (Re_p and Re_c) as shown in (3) and (4)

$$Re_p = \frac{\rho v d_p}{\mu} \tag{3}$$

$$Re_c = \frac{\rho v D_e}{\mu} \tag{4}$$

taking (3)/(4)

$$\frac{Re_p}{Re_c} = \frac{d_p}{D_e}. \tag{5}$$

The particle-to-bed-diameter ratio is tabulated in Table IV. It can be seen that the particle-diameter-to-bed-diameter ratio in gasifier is always larger than the combustor.

TABLE IV. PARTICLE-TO-BED-DIAMETER RATIO

Compartment	**Gasifier**	**Combustor**
Mean particle size, d_p (mm)	**d_p/D_e (mm/m)**	
0.196	0.76	0.47
0.272	1.05	0.66
0.341	1.33	0.83
0.395	1.54	0.96

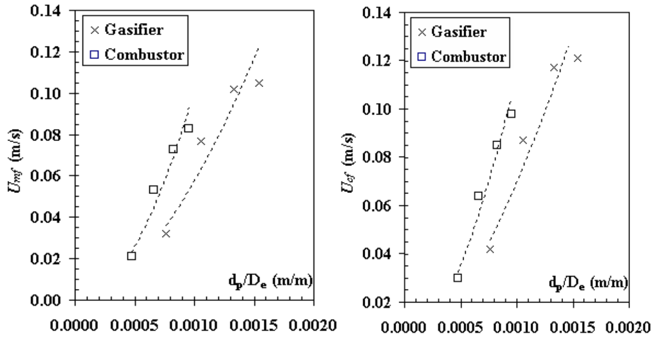

Figure 7. Characteristic velocities vs. dimensionless diameter

The characteristic velocity vs. particle-to-bed-diameter profiles are shown in Fig. 7. These relationships can be described as $U^i_{mf} = f\left(\frac{d_p}{D_{e,i}}\right)$ and $U^i_{cf} = f\left(\frac{d_p}{D_{e,i}}\right)$ respectively where 'i' represents different column diameter. The system (whole CFBG) U_{mf} and U_{cf} increases roughly in accordance to $(d_p/D_e)^{1.35}$ and $(d_p/D_e)^{1.18}$ respectively. Similarly, J.F. Frantz [13] reported that U_{mf} increases with the particle-to-bed diameter ratio according to $(d_p/D_e)^{\alpha}$ where α>0.

D. U_{mf} and U_{cf} Values Comparison With Correlations

Most of the characteristic velocity correlations for multicomponent system are based on binary mixtures (requiring input parameters of the two components only) [5]-[6], thus are not used for the present wide size distributed powder. Instead, for the comparison of U_{mf}, the notable Wen and Yu correlation [14], Thonglimp et al. [15] and Reina et al. [16] are selected. On the other hand, the U_{cf} values are compared with correlations from Gauthier et al. [9], Vaid and Gupta [17] and Mourad et al. [18].

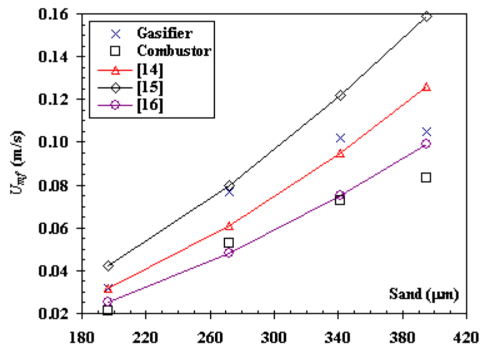

Figure 8. U_{mf} comparison with selected correlations

Fig. 8 shows the comparison of experimental U_{mf} values with selected correlations. It can be seen that only Wen and Yu correlation yields the closest predictions for both compartments, providing nearly average values for the upper (gasifier) and lower (combustor) U_{mf} values. On the other hand, Fig. 9 shows the comparison of experimental U_{cf} values with selected correlations. All the correlations overestimate, although correlation [9] provides the closest prediction to the U_{cf} values.

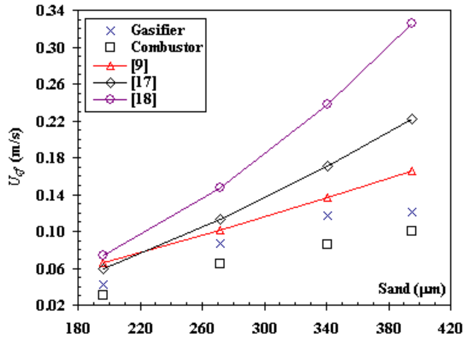

Figure 9. U_{cf} comparison with selected correlations

IV. Conclusion

Based on the experiments carried out, it can be concluded that CFBG characteristic velocities are closely resemble to those reported in cylindrical fluidized bed of laboratory scale. The U_{mf} and U_{cf} values are affected by the particle size distribution and bed geometry. The former leads to $U_{cf} > U_{mf}$ and U_{cf}/U_{mf} ratio is approximately constant (except for 196µm) for gasifier and combustor respectively, while the latter leads to $(U_{cf}, U_{mf})_{gasifier} > (U_{cf}, U_{mf})_{combustor}$ due to cumulative effects of the bed diameter and particle-to-bed diameter ratio. None of the selected correlations predicted well in both compartments as they do not include these effects. Meanwhile, the governing parameters can be minimized when utilizing larger sand size as bed material, hence avoiding physical modification on the vessel.

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