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Effect of Novel Swirl Distributor Plate on Hydrodynamics of Fluidized Bed Gasifier

I. E. Afrooz*, D. L. Chuan Ching

Department of Fundamental and Applied Sciences, University Technology PETRONAS, Malaysia

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ABSTRACT

The main advantage of a fluidized bed is its capability in excellent gas-fuel mixing. However, due to the lacks of gas radial momentum, its lateral mixing of gas-solid is not adequate. Therefore, this research is focused on fluidized bed hydrodynamics enhancement using the modified gas distributor plate design. For getting an optimum fluidized bed hydrodynamics prediction, three different classical ANSYS Fluent drag models, namely Wen-Yu, Syamlal O'Brien, and Gidaspow are examined first. Afterward, a novel distributor plate called swirl distributor plate (SDP) is proposed in order to enhance the gas-fuel mixing in vertical and radial directions. In terms of simulation approach, results were presented and compared with the experimental data. It has been observed that better hydrodynamics prediction is achieved by Syamlal O'Brien drag model. The effect of SDP on gas-solid mixing was then studied numerically and compared with conventional distributor plate (CDP). Compared with CDP, better gas-solid mixing was found while the SDP was used. As a final point, gasification test was conducted in a lab scale system in order to study and compare the composition of produced syngas using both distributor plates. Based on the gasification results, SDP leads to promotion of Hydrogen and Carbon monoxide by 34.85 and 65.92 percent, respectively.

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NOMENCLATURE

		Greek Symbols	
g	*Gas*	ρ	Density (kg/m^3)
s	Solid	μ	Viscosity (kg/m.s)
d	Diameter (m)	τ	Relaxation time (s)
K	Exchange coefficient	ε	Volume fraction
U	Superficial velocity		
Re	Reynolds number		

1. INTRODUCTION

The difficulty to distribute fluidizing medium into a fluidized bed homogenously has led to innovation in air supply and distributor design such as controlling the bubble size with flow pulsation distributor [1], vibrated beds [2], sound assisted beds [3], spiral distributor [4], and swirling distributor [5]. In 1998, Lakshmanan and Dodson [6] introduced a Torbed gas-solid reactor which can be considered as the first demonstration of swirl distributor. Considering Lakshmanan and Dodson distributor, high heat and mass-transfer rates has been reported. Köksal et al. [1] proposed Moving Double Plate Distributor (MDPD) to improve the fluidization characteristics by controlling the size of the bubbles. It was stated that using pulsed flow reduced the bubble size significantly due to bubble splitting in the pulsed flow.

The swirling fluidized bed hydrodynamic characteristics was studied by Sreenivasan and Raghavan [4] using an spiral distributor plate. Authors of the article stated that swirling motion of solid particles can be achieved by using spiral distributor plate. However, the performance of spiral distributor plate for larger radii has not been approved yet. Wilde and Broqueville [7] tested the influence of radial gas injection on the hydrodynamics of the fluidized bed reactor (FBR). As it was mentioned by the authors, the problem of slugging, channeling and an irregular distribution of the gas may

*Corresponding Author Email: *imaneslami@hotmail.com* (I. E. Afrooz)

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occur. Besides the complexity of the multiple gas injection points design, the problem of solids losing via the chimney make this design to some extent ineffective. Therefore, applying an optimization design is essential.

Sobrino et al. [8] investigated the performance of a novel rotating distributor plate on fluidized bed hydrodynamics. It was found that the rotating distributor plate promotes radial dispersion of the particles. However, the improvement in the quality of fluidization was disappeared when the gas velocity became bigger than the tangential velocity. Sobrino et al. [9] studied the effect of bubble caps and perforated plate distributors on the flow configuration near the bottom area of a turbulent fluidized bed. It was concluded that the perforated plate increased the rate of solids transferred from the dense bed to the riser and it began at a lower superficial gas velocity. Moreover, it was found that bubble-cap distributors result in higher solids density at the bottom of the bed. Howver, in terms of voidage, a regular radial structure was found.

In 2015, Aworinde et al. [10] proposed a swirling flow nozzle for enhancing fluidization quality of a fluidized bed. Compared to the simple nozzle, better gas–solid contact was observed whit swirling flow nozzle. Although the above research shows the remarkable results for the swirling flow nozzle, but the investigation has been done only on a single nozzle. Therefore, more research needs to be done to examine the performance of multiple swirling flow nozzles in a distributor plate. In 2011, Xuliang et al. [11] studied the effects of sintered metal distributor (SMD) on the hydrodynamics of fluidized bed. It was found that, SMD results in a more stable fluidization.

The influence of a novel multi-vortex (MV) on a 2D fluidized bed’s gas axial dispersion, bubble size, and mass transfer was investigated by Brink et al. [12]. MV distributor resulted in 50% greater increase in mass transfer coefficient. Considering the overall FBR performance improvement while using MV distributor, the authors stated that the bubble sizes is opposed to the mechanism of mass transfer. However, no quantification was provided to support the argument validity. Moreover, more research is required to investigate the effect of this kind of distributor in a three-dimensional domain.

In a nutshell, although some progress has been made toward the fluidized bed hydrodynamics improvements, the gas-solid lateral mixing enhancement still remains controversial and should be investigated more. In this paper, a novel swirl distributor plate (SDP) is proposed to enhance fluid dynamic characteristics of the fluidiezed bed as well as petcoke gasification over conventional distributor plate (CDP).

Earlier to fabricate the cold model, the numerical model has been developed using ANSYS Fluent 15. The data acquired using numerical techniques have been applied to optimize the cold model design. Furthermore, the cold and hot flow models were fabricated and the effect of two different distributor plates (perforated and swirl) were examined by observing the percentage of combustible gases produced in the process. Finally, the carbon conversion efficiency and cold gas efficiency were then calculated to understand the efficiency of the system.

This study is important because it shows whether the change in the flow profile from two dimensions in CDP (2D) to three dimensions (3D) in SDP influence the hydrodynamic of the fluidized bed by enhancing gas/solid mixing, improving heat transfer and pressure drop and increasing the particle residence time.

In the next section, the methodology used in modelling and fabricating of the fluidized bed reactor is discussed. The numerical and experimental results are presented and compared in the results section. A summary of the results obtained and general conclusions are discussed in the last section.

2. METHODS

2. 1. Simulation Setup Figure 1 demonstrates the schematic view of fluidized bed used in this study. As it can be seen from the Figure, the fluidized bed is a half meter cylinder with the bed diameter of 0.0762 m. Pressure outlet and velocity inlet boundary conditions are employed at the top of the riser and at the bottom of the bed, respectively (Figure 1a). The solid model is then divided into finite elements by hexahedral mesh technique because it gives the high accuracy of the solution and reduces the computational time (Figure 1b).

One of the important factor that shows the mesh quality is skewness. After analyzing the mesh quality by the mesh metric, the average skewness of 0.19 was detected which is far below the critical value of 0.8. A

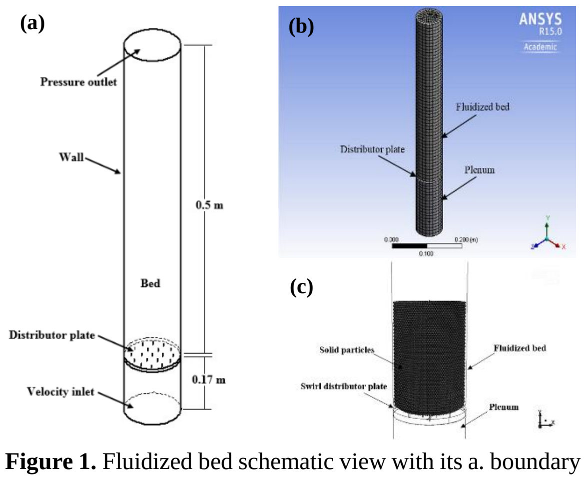


Figure 1. Fluidized bed schematic view with its a. boundary condition, b. grid construction, and c. loaded solid particles

total 94380 spherical shape solid particles with a total weight of 0.5 Kg and diameter of 500 µm were placed in the fluidized bed (Figure 1c). To this end, codes to generate the possible places where particles can be placed was written using MATLAB programming software. Then, the particles were distributed inside the fluidized bed using file injection method. The other simulation parameters is tabulated in Table 1.

2. 2. Drag Models In this work, Wen-Yu, Syamlal O'Brien, and Gidaspow drag models have been used and compared. According to the Syamlal O'Brien model [13], the exchange coefficient between fluid and solid (K_{sg}) is defined as follows:

$$K_{\mathrm{sg}} = \frac{\varepsilon_s \varepsilon_g \rho_g C_D R_e}{24 \tau_s v_{r,s}^2} \tag{1}$$

where,

$$C_D = \left[0.63 + \frac{4.8}{\sqrt{Re_s / v_{r,s}}}\right]^2 \quad [14] \tag{2}$$

$$v_{r,s} = 0.5[A - 0.06Re_s + \sqrt{(0.06Re_s)^2 + 0.12Re_s(2B - A) + A^2}] \quad [15] \tag{3}$$

with

$$A = \varepsilon_g^{4.14}$$

$$\text{and} \begin{cases} B = 0.8\varepsilon_g^{1.28} & \text{for } \varepsilon_g \leq 0.85 \\ B = \varepsilon_g^{2.65} & \text{for } \varepsilon_g > 0.85 \end{cases} \tag{4}$$

and relative Reynolds number is

$$Re_s = \frac{\rho_g d_s |\vec{v}_{\mathrm{s}} - \vec{v}_{\mathrm{g}}|}{\mu_g} \quad [16] \tag{5}$$

The particle relaxation time τ_s is calculated via,

$$\tau_s = \frac{\rho_s d_s^2}{18\mu_g} \quad [16] \tag{6}$$

where d_s is the particle diameter.

Considering Wen-Yu drag model, the fluid-solid exchange coefficient is defined as follows:

$$K_{\mathrm{sg}} = \frac{3}{4} C_D \frac{\varepsilon_s \varepsilon_g \rho_g |\vec{v}_{\mathrm{s}} - \vec{v}_{\mathrm{g}}|}{d_s} \varepsilon_g^{-2.65} \tag{7}$$

TABLE 1. Simulation condition used in the research

Simulation Parameters	**Value**
Bed diameter (mm)	76.2
Orifice diameter, d_{or} (mm)	1
Gas density, ρ_g (Kg/m^3)	1.225
Gas viscosity, µ (10-5 kg/ms)	1.79
Solid density, ρ_s (Kg/m^3)	2608
Solid diameter, d_s (µm)	500
Superficial velocity, U (m/s)	0.46

where

$$C_D = \frac{24}{\varepsilon_g Re_s}\left[1 + 0.15(\varepsilon_g Re_s)^{0.687}\right] \tag{8}$$

and Re_s is defined by Equation (5).

Gidaspow fluid-solid drag model [17] is a combination of the Wen and Yu model [18] and the Ergun equation [19]. Based on this model, the exchange coefficient between gas and solid (K_{sg}) is of the following form:

when $\varepsilon_g > 0.8$

$$K_{\mathrm{sg}} = \frac{3}{4} C_D \frac{\varepsilon_s \varepsilon_g \rho_g |\vec{v}_{\mathrm{s}} - \vec{v}_{\mathrm{g}}|}{d_s} \varepsilon_g^{-2.65} \tag{9}$$

where

$$C_D = \frac{24}{\varepsilon_g Re_s}\left[1 + 0.15(\varepsilon_g Re_s)^{0.687}\right] \tag{10}$$

$$Re_s = \frac{\varepsilon_g \rho_g d_p |\vec{v}_{\mathrm{s}} - \vec{v}_{\mathrm{g}}|}{\mu_g} \tag{11}$$

when $\varepsilon_g \leq 0.8$,

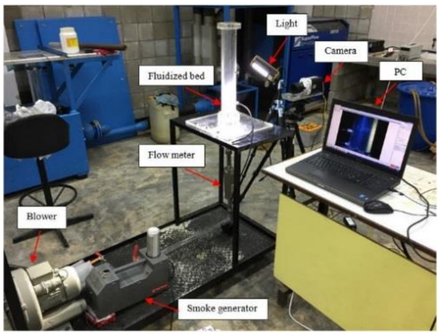
$$K_{\mathrm{sg}} = 150 \frac{\varepsilon_s(1-\varepsilon_g)\mu_g}{\varepsilon_g d_s^2} + 1.75 \frac{\rho_g \varepsilon_s |\vec{v}_{\mathrm{s}} - \vec{v}_{\mathrm{g}}|}{d_s} \tag{12}$$

2. 3. Experimental Setup There are two types of fluidized beds used in this study, namely cold and hot flow models. The cold flow model, as its name says, is used to study the hydrodynamics of the fluidized bed at ambient temperature and pressure (Figure 2). It is mainly comprised of a fluidized bed, a blower, and a smoke generator.

The cold model is a half meter Plexiglas column with 0.0762 m diameter. One kg sand with the size of 500 µm and density of 2608 kg/m^3 is placed in the bed column. This amount of sand occupies 0.1 m height of the bed. In order to study gas-solid mixing hydrodynamics, air is injected into the system using a blower and through a flow meter and distributor plate. Then, the flow structure was characterized using a high speed camera with the photography rate of 1000 frame per second and a source of distributed light.

The hot flow model (Figure 3) includes the combustion process and the operating temperature of the reactor is 800-900°C. Therefore, a 304 stainless steel pipe with the melting point of 1450°C would be the right choice for the hot model body structure. This pipe is inexpensive and has a good resistance to oxidation for intermittent use, up to 925 °C.

Main components of the unit are blower, screw feeder, fluidized bed, cyclone and gas analyzer for data acquisition. The fluidized bed contains half Kg of sand. The preheated air enters the reactor from the bottom and fluidize the bed materials (sand) while feedstock (Petcoke) enters the fluidized bed from a side and mixes with the oxygen. After going through the gasification



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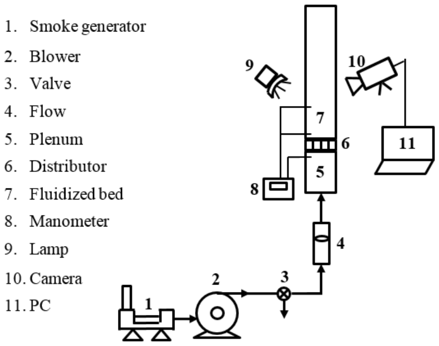

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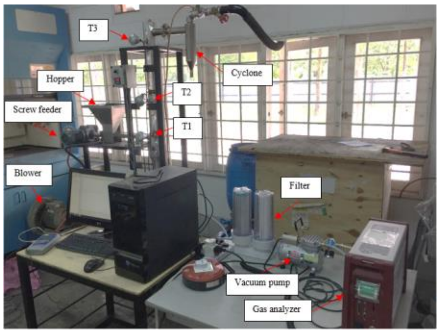
Figure 2. Bubbling fluidized bed setup a. cold flow model and b. its schematic view



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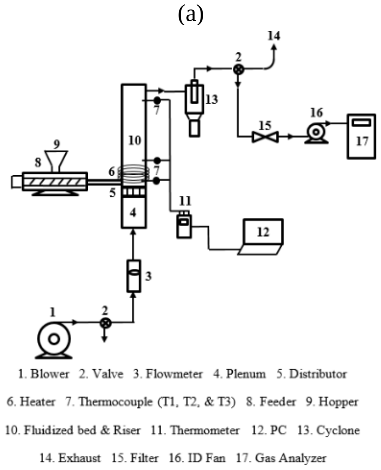


(b)

Figure 3. Bubbling fluidized bed setup a. hot flow model and b. its schematic view

reaction, the product gases (syngas) are discharged from the exhaust and analyzed using gas analyzer to determine the composition of each component in the syngas.

Tests were made using two different distributor plates (Figure 4) in order to study their effects on fluidized bed hydrodynamics and syngas compositions using cold and hot flow models, respectively.

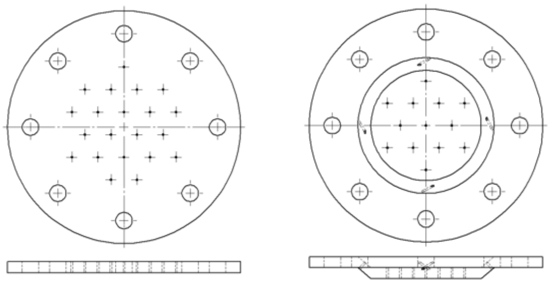
3. RESULTS

In this section Syamlal O'Brien, Wen-Yu, and Gidaspow drag models are evaluated first to get a drag model with the optimum fluidized bed hydrodynamics prediction. Next, the results of the effect of the distributor plates on hydrodynamics of the fluidized bed using cold flow model is discussed. Afterwards, the gasification results of petcoke using hot flow model and different distributor plates are presented.

3. 1. Drag Model Evaluation The flow structure of the fluidized bed simulated by the above mentioned drag models is presented in Figure 5. As can be inferred from the Figure, the flow pattern is almost same for all drag models till the eruption of the first bubble. Afterward, the difference in flow patterns can be observed.

In order to assess the accuracy of the drag models in predicting the fluidized bed hydrodynamics, the simulation results of bubble movement and bed expansion were compared with those of experimental data. In terms of bubble movement, good agreement was observed between the experiment test results and the results obtained by Syamlal O'Brien drag model (Figure 6).

Figure 7 depicts a comparison of bed expansion's simulation and experimental results at three different time recording of 0.0, 0.54, and 1.0 second. The bed height at time zero and 0.54 second refer to the initial bed height which is 0.1 meter (10 cm) and the time after bursting the first bubble, respectively. By looking at the figure, the average bed expansion of 2.5 cm can be observed. However, the particles were thrown up to the height of the 20 cm.



(a) (b)

Figure 4. Schematic view of (a) CDP and (b) SDP

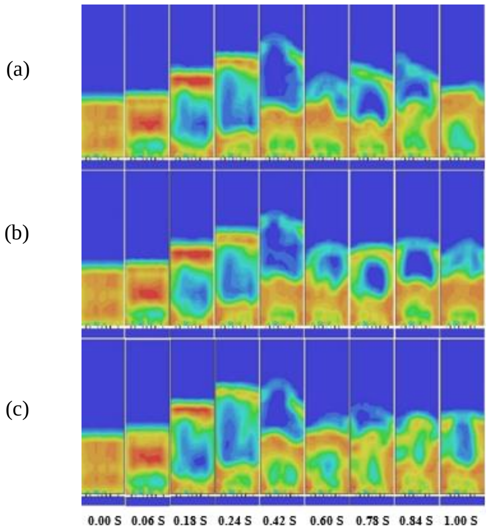


Figure 5. Solid volume fraction's contour plots with, a. Gidaspow, b. Syamlal O'Brien, and c. Wen-Yu drag models

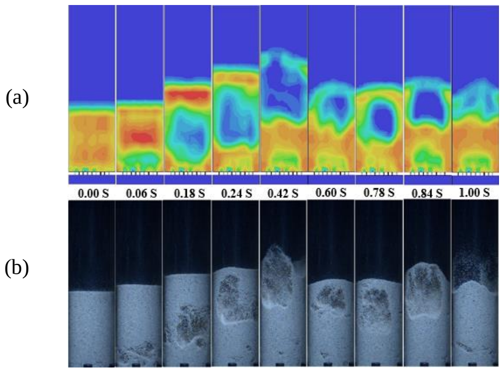


Figure 6. Comparison between a. Syamlal O'Brien drag model simulation results and b. experimental results

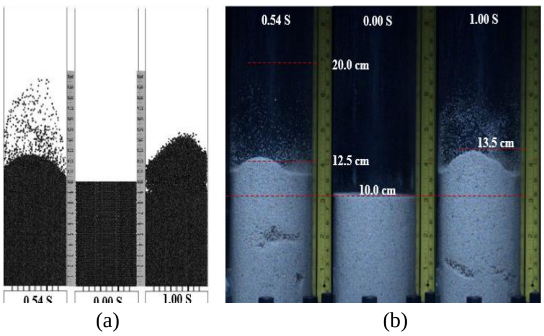


Figure 7. Comparison between a. Syamlal O'Brien drag model simulation results and b. experimental results

Figure 8 has provided to clarify the method in measuring the initial bed height. As it can be seen from the photo, the initial bed height is measured from the top surface of the distributor plate (Figure 8a). However, the ruler was then dislocated to the top of the flange to have a better reading (Figure 8b). Therefore, by doing this, the value of 7.5 cm indicates the value of 10 cm. This 2.5 cm difference is the distance between the top of distributor plate and the flange. Considering that, the initial height of the bed at time zero is 10 cm (Figure 7b middle). Similar trend can be observed by comparison of simulated bed expansion with experiment data. The height in which the solid particles thrown up is also detected to be the same.

3. 2. Hydrodynamics Investigation

The overall behavior of particle movements can be studied using the time-averaged distributions of solid velocity. To this end, the vector diagram of instantaneous particle velocity after 1 second from the initial state for both CDP and novel SDP is plotted and compared in Figure 9.

As can be clearly seen from the graph, solid particles movements in the bed with CDP are upward, downward, and inner circulation in the center, near the sidewalls, and throughout the bed, respectively. However, particles

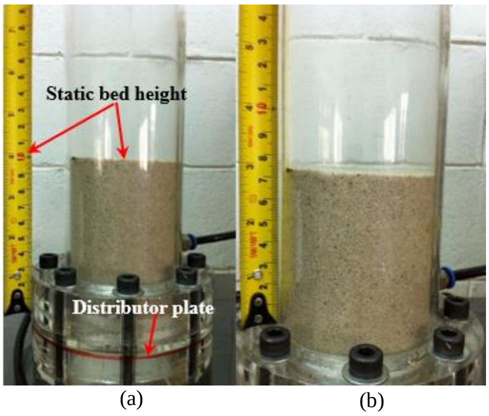


Figure 8. Initial bed height a. from distributor plate b. from flange top

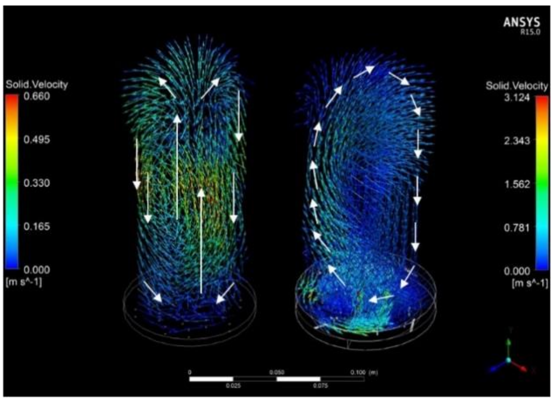


Figure 9. Solid velocity distribution pattern of fluidized bed with CDP (Left) and SDP (Right). White arrows display the straight and swirl flow directions for CDP and SDP, respectively

movements following different pattern while using the SDP. The solid particles are swirling upward from one side and swirling downward from the other side of the bed due to the radial momentum of the gas phase. It is worthwhile to point out that due to the swirl movement, the radial mixing of the solid particles is increased significantly.

Another important observation about the particle velocity distribution graph is that the highest and the lowest particles velocities appeared in the middle and the bottom of the bed having CDP. Therefore, the possibility of the dead zones formation at the bottom of the fluidized bed is considerably high. In contrast, using the novel SDP caused the intense gas radial momentum near to the bed bottom. Therefore, the highest solid particles velocities can be seen in that region which decreases the possibility of dead zone formation. Another point worth mentioning about SDP is that it increases the particle velocity to the magnitude of 5 compare with CDP. This rise in the particle velocity increases the mixing rate and as a result the rate of heat transfer increases. The particle tracking technique (Figure 10) was also used to track one individual particle during its 3 seconds traveling inside the fluidized bed.

As it was expected, using CDP, the fluidized particle travels upward and downward in the middle and next to the wall of the bed, respectively. However, the motion trajectory of the particle fluidized with the SDP is swirl path.

3. 3. Gasification Analysis The results of syngas compositions for both distributor plates is presented in Figure 11. Generally, the results show that the syngas produced while SDP was used is relatively richer in carbon monoxide, hydrogen and methane than that of CDP, and therefore SDP results in better syngas composition. On the other hand, it can be seen that the syngas contains also large amounts of carbon dioxide, however, considering SDP, the CO_2 emission reduces as the amount of syngas increases.

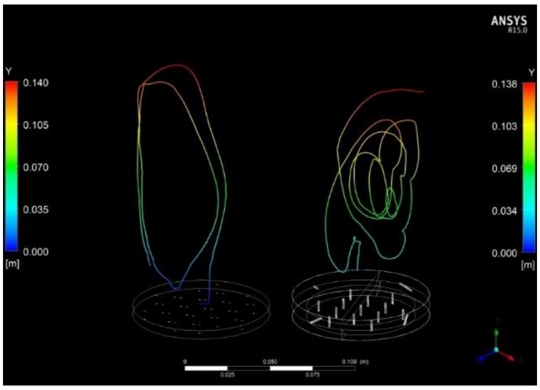
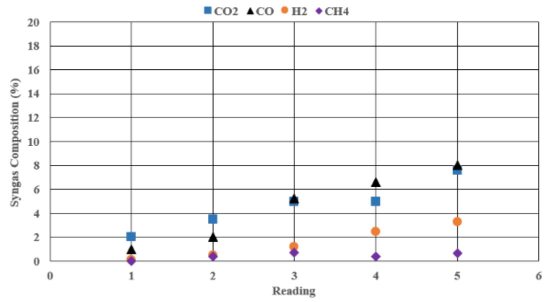


Figure 10. Particle track for both distributor plate at time t = 3s: Up and Down for CDP (left) and Swirl for SDP (right)



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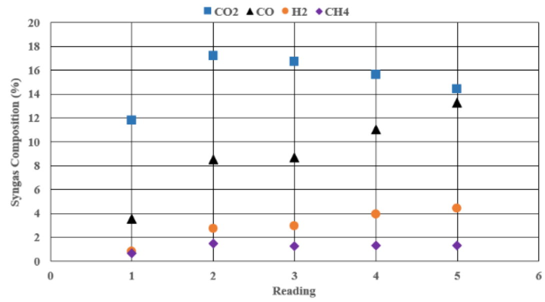

(b)

Figure 11. Syngas composition at different reading time while a. CDP, and b. SDP were used. SDP resulted in syngas with higher Carbon monoxide and Hydrogen compositions

Cold gas and carbon conversion efficiencies of the gasifier system were calculated and results are shown in Table 2. The results show that more carbon was converted to the gases when gasifier equipped with SDP. Moreover, high cold gas efficiency indicates an enhancement in gasifier performance when SDP was used. It is worth to mention that the poor carbon conversion is resulted from low residence time of the char particles [20]. Therefore, it is without doubt that the particles swirl motion, imposed by SDP, improves the particles residence time.

TABLE 2. Comparison of carbon conversion and cold gas efficiency.

	Carbon conversion efficiency (%)	Cold gas efficiency (%)
CDP	13.77	6.79
SDP	29.67	13.42

4. CONCLUSIONS

In this paper, the performance of the novel swirl distributor plate on hydrodynamics of a fluidized bed and as a result on syngas production was verified numerically and experimentally. In terms of numerical investigation, three existing drag models evaluated first and as a result the Syamlal O'Brien drag model has been found to possess better prediction accuracy and applicability.

Simulation results confirmed that the proposed swirl distributor plate promotes gas-fuel mixing and consequently production of H_2, CH_4 and CO gases were improved by 34.85, 100.00, and 65.92 percent, respectively.

5. ACKNOWLEDGMENTS

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Computational Fluid Dynamics

چکیده

مهمترین مزیت رکتورهای بستر سیّال قابلیت آنها در مخلوط کردن ترکیب سوخت با گاز می‌باشد. اما به دلیل فقدان حرکت شعاعی، میزان مخلوط شدن در جهت افقی بسیار کم می‌باشد. بنابر این، هدف از این تحقیق، بهینه کردن عملکرد صفحه توزیع کننده گاز در جهت تقویت هیدرودینامیک رکتورهای بستر سیّال می‌باشد. در ابتدا و در جهت پیدا کردن بهترین مدل برای پیشبینی دقیق هیدرودینامیک این رکتور ها، سه دراگ مدل کلاسیک موجود در نرم افزار انسیس به نام های Wen-Yu, Syamlal O’Brien و Gidaspow مورد بررسی قرار میگیرند. با مقایسه دادهای بدست آمده از شبیه سازی با دادهای عملی، بهترین پیشبینی با استفاده از دراگ مدل Syamlal O’Brien نتیجه شده است. در ادامه، یک صفحه توزیع کننده گاز به نام صفحه توزیع کننده چرخشی (SDP) به منظور افزایش مخلوط شدن ترکیب سوخت با گاز در هردو جهت عمودی و شعاعی ارائه شده است. به دنبال آن، اثر صفحه توزیع کننده چرخشی (SDP) و معمولی (CDP) بر کیفیت مخلوط سوخت و گاز شبیه سازی و مقایسه شده است. نتایج نشان میدهد که SDP باعث بهبود در مخلوط سوخت و گاز میشود. و در انتها، عملکرد هردو صفحه توزیع کننده بر تولید گازهای مفید (Syngas) مورد آزمایش قرار گرفته است. بر اساس نتایج بدست آماده، SDP منجر به ارتقاع هیدروژن و منوکسید کربن به میزان ٣٤.٨٥ و ٦٥.٩٢ درصد میشود.

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