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Research on the Influence of the Scaling-up Process on the Flow Structure and Two-Phase Distribution of Gas–Solid Fluidized Beds

Zhonglin Gao ^{1,*}, Jie Wei ¹, Hongyang Wang ¹  and Chenyang Zhou ²

¹ School of Materials Science and Engineering, Anhui University of Science and Technology, Huainan 232001, China; weijie05312024@163.com (J.W.); hywang3@163.com (H.W.)

² School of Chemical Engineering and Technology, China University of Mining and Technology, Xuzhou 221116, China; zhoucy@cumt.edu.cn

* Correspondence: zlgao2022@163.com

Abstract: In the continuous scaling-up process of the separating system, a mechanism exists that transforms the behavior of the flow field, resulting in deviations from the original model and conclusions. The paper examined the effects of the scale up of a fluidized bed by CFD. It was observed that increasing the diameter reduces the amplitude of axial density fluctuations. Similarly, increasing the static height increases the amplitude. Moreover, increasing the static bed height enhances the visibility of the cyclic flow structure of gas and solid phases. The flow structure in large bed diameters is disrupted. The impact of changing the bed diameter on bed density is more significant than the static height. As the bed diameter increases, the bubble disturbance decreases and the aggregation phase gradually disappears while the proportion of the emulsified phase keeps increasing. This study will guide and assist in the future application of separated fluidized beds in industry.

Keywords: TFM; scale-up; fluidized bed; density; two-phase distribution



Citation: Gao, Z.; Wei, J.; Wang, H.; Zhou, C. Research on the Influence of the Scaling-up Process on the Flow Structure and Two-Phase Distribution of Gas–Solid Fluidized Beds. *Separations* **2024**, *11*, 202. <https://doi.org/10.3390/separations11070202>

Academic Editor: Mingheng Li

Received: 2 June 2024

Revised: 22 June 2024

Accepted: 25 June 2024

Published: 28 June 2024



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1. Introduction

At present, research on the fluidization and separation characteristics of fluidized beds is limited to technical theories and conditions and is conducted at the laboratory level. The structure formed by multi-component particles is complex, and there is a mechanism of flow field behavior transformation during the scaling-up process, which leads to deviations in the theoretical prediction model and experimental conclusions [1]. In addition, when gas–solid separation fluidized beds are scaled up, the size of bubbles increases significantly, particle back-mixing becomes more severe, and abnormal fluidization phenomena like channel flow and agglomeration occur more quickly in the bed. As a result, the bed density becomes uneven, and the efficiency of mineral particle separation deteriorates. These issues are commonly known as the system’s scaling-up effect [2].

Circulation, flow, and contacting patterns in large reactors differ from those in small-scale units [3]. In the pilot reactor, the bubble size reached the bed diameter, causing slugging. In large commercial reactors, bubbles could grow continuously, and their rise velocity also increased proportionally [4]. The slug rise velocity is slower than the bubble rise velocity [5]. De Groot et al. [6–8] discovered that bed expansion depends on bed diameter and is affected by wall effects. Contact with the reactor wall may impact the interaction between particles and gas in the bed. When enlarging the gas–solid fluidized bed, it is necessary to consider the impact of the reactor diameter on fluid dynamics [9,10]. Frye et al. [11] studied how bed diameter affects the chemical conversion of ozone decomposition reactions. The reaction rate was observed to decrease gradually as the bed diameter increased, with three sizes examined ($R = 0.051$ m, 0.203 m, and 0.762 m). In addition, Fu [12] observed that the minimum fluidization velocity of B-type particles increases

gradually as the bed height increases. Scaling laws for fluidized beds have been developed by Glicksman [13] and Horio [14]. The purpose of these laws is to serve as a reference for creating small-scale, cold flow models that exhibit the same hydrodynamic behavior as large, hot commercial units.

The fluidized bed's deviation from the theoretical model and experimental conclusions may be attributed to scale issues. Therefore, accurately revealing the scaling-up effect during the continuous scaling-up process of the separation fluidized bed is crucial for restricting its industrial scaling-up.

Experimental research is frequently restricted by measurement methods and testing conditions, which leads to most studies being confined to macroscopic scales. Obtaining all possible microscale information in laboratory- and industrial-scale research is not possible. Simple experimental methods are not effective in identifying the complex structure of fluid-particle systems. Numerical simulation has become an important method for studying gas–solid two-phase interaction in this context. It can provide theoretical guidance for regulating fluidized beds at micro- and macroscales [15–18].

Currently, there are two main models used for simulating gas–solid fluidized beds: the Euler–Euler (EE) model and the Euler–Lagrange (EL) model. EL models usually consider the solid phase as a separate entity called the Discrete Element Method (DEM), while the gas phase is treated as a continuous phase. The EE model is alternatively referred to as the Two-Fluid Model (TFM) [19–22]. The Navier–Stokes equation is used to model the gas and solid phases as interpenetrating continuums. TFM has a major benefit in that it can handle a large number of particles without the need to track the trajectory of each individual particle. CFD-DEM is also a commonly used simulation method, but it also has limitations such as high computational complexity, dependence on particle material models, sensitivity to time steps, and sensitivity to boundary conditions [23–25]. ANSYS Fluent achieves the best convergence speed and accuracy by utilizing multiple solving methods and multi-grid accelerated convergence techniques [26–28].

In the traditional chemical industry, researchers have scaled up reactors based on proposed scaling methods and achieved good research results [29,30]. But, at the current stage, there is no clear conclusion regarding gas–solid separation fluidized beds; therefore, a fixed proportion cannot be used for scaling up. In this paper, we gradually increase the bed diameter and height to simulate the scaling-up process of the system and hoped to reveal the scaling-up effect of fluidized dry separating machines. This study utilized the ANSYS Fluent 2021R1 software with Two-Fluid Model in order to analyze the variations in axial and radial density fluctuations, gas–solid two-phase flow velocity, and two-phase distribution during the scale-up process of gas–solid fluidized beds.

2. Numerical Method

The simulation parameters are provided in Table 1. The bed size was configured in a two-dimensional view. Gas was introduced from the bottom, with an initial particle volume fraction set at 0.6. The wall condition was maintained as a no-slip condition, and additional details are presented in Table 1. The simulation process sets various static bed heights and diameters as follows: 20 cm, 30 cm, 50 cm, and 70 cm for bed heights and 10 cm, 15 cm, and 38 cm for bed diameters, as shown in Figure 1. The fluidization regime belongs to bubbling of fluidized bed. All control equations are listed in the Appendix A.

Table 1. Parameter set for the simulation.

Property	Value
Gas velocity	0.2 m/s
Gas density	1.29 kg/m ³
Gas viscosity	1.8×10^{-5} Pa·s
Particle diameter	0.0002 m
Particle density	4500 kg/m ³
Restitution of coefficient	0.9
Initial particle volume fraction	0.6
Time step	0.0001 s
Drag model	Gidaspow
Viscous model	laminar

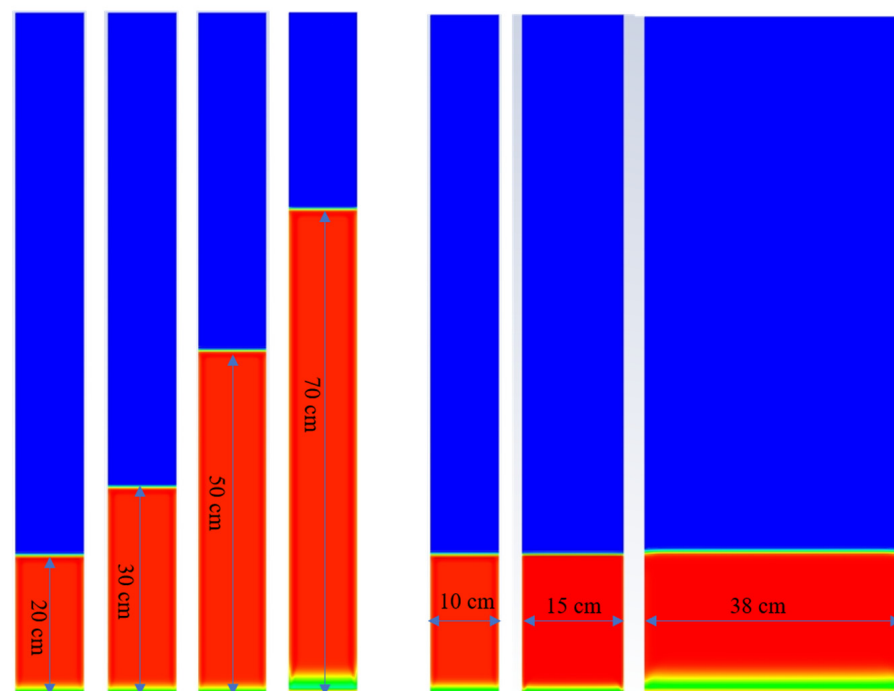


Figure 1. Different fluidized bed sizes.

3. Results and Discussion

3.1. Grid Independence

The parameter of grid size has a significant impact on the accuracy and efficiency of the simulation process. Therefore, before formal discussion, this section will first discuss the grid independence of the simulation results. We use four grid sizes (3 ds, 5 ds, 10 ds, 20 ds), where dp is the particle size. Gas velocity $U_g = 0.2$ m/s, static bed height $H_s = 20$ mm. Figure 2 shows the variation of static pressure along the axial height simulated under different grid sizes, with a statistical time of 10 s. It can be seen that the distribution trend of static pressure for grid sizes of 10 times and 20 times particle sizes is basically consistent. Similarly, conducting grid independence checks on fluidized beds of other diameters ultimately yields the optimal grid size. Considering the calculation time factor, a grid with a particle size of 10 times will be used as the standard for all simulation calculations in subsequent discussions.

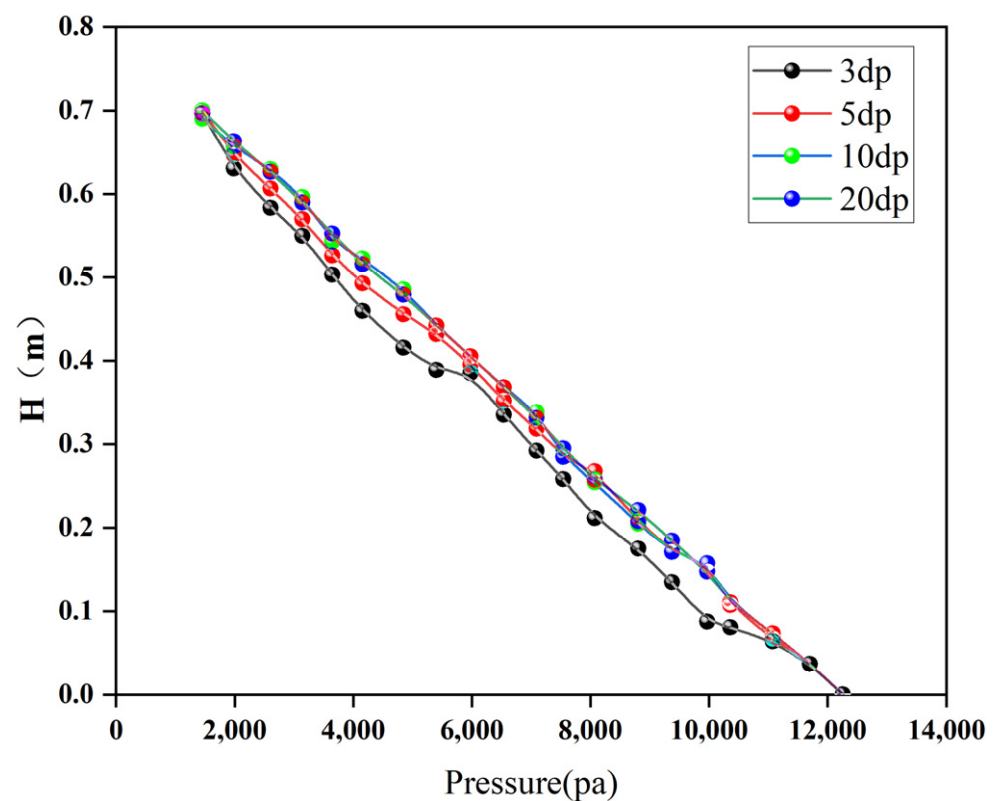


Figure 2. The influence of grid size on the axial pressure ($R = 15$ cm).

3.2. Changes in Axial Density of the Bed

Figure 3 shows the axial density variation of the bed at position $2/D$ across various bed diameters ($D:10, 15, 38$) and bed heights ($H:20, 30, 50, 70$). From the graph, it is evident that when the bed height is constant, increasing the bed diameter leads to a decrease in the amplitude of axial density fluctuation in the bed. Similarly, when the bed diameter remains constant, increasing the static bed height results in a gradual increase in the amplitude of axial density fluctuation in the bed. Based on Figure 4, it can be observed that the bed diameter and static bed height have minimal impact on the axial average density. Under the same static bed height conditions, the average bed density increases, and the standard deviation of density fluctuations decreases as the bed diameter increases. This is because, when combined with Figure 5, the smaller the bed diameter, the bubble can grow and approach the bed diameter more easily. This leads to severe density fluctuations and a decrease in the average bed density. On the contrary, as the bed diameter increases, the impact of bubble motion on global density fluctuations decreases, and the average bed density increases. Furthermore, when considering the same bed diameter, the influence of static bed height on the average axial density of the bed is not substantial. However, it does have a notable effect on the fluctuation of axial density. Specifically, as the static bed height increases, the standard deviation of axial density fluctuation of the bed also increases. The reason is that as the static bed height increases, the growth of bubbles becomes closer to sufficient and has a greater impact on the axial density of the bed layer.

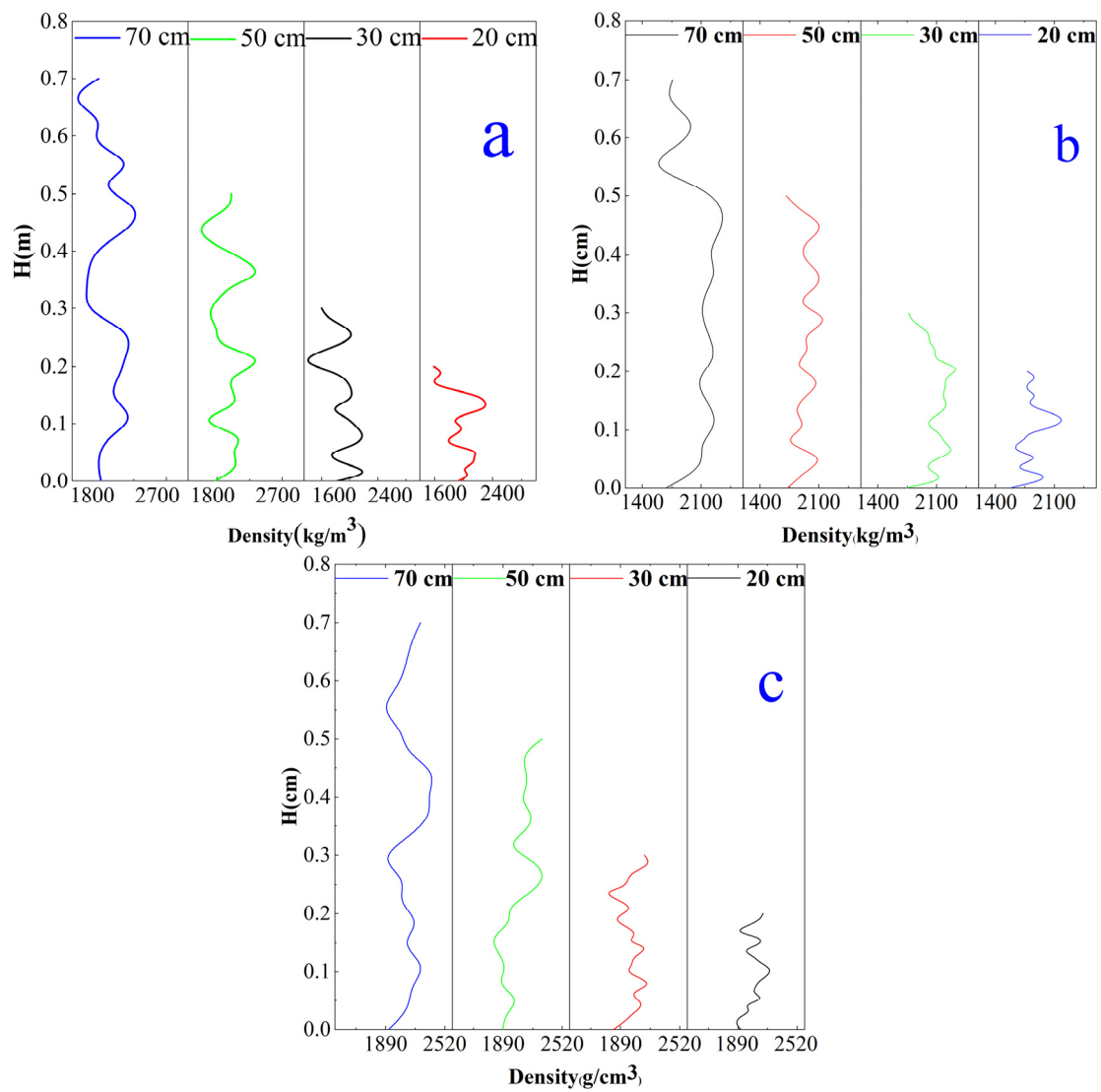


Figure 3. Changes in axial density of the bed ((a): $D = 10$ cm, (b): $D = 15$ cm, (c): $D = 38$ cm).

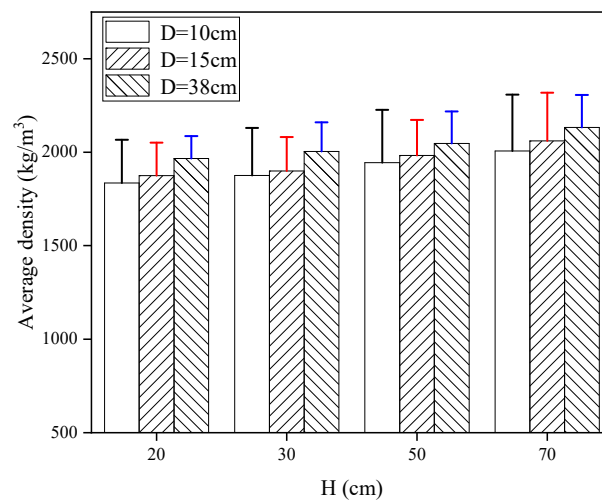


Figure 4. Average density variation with error bar.

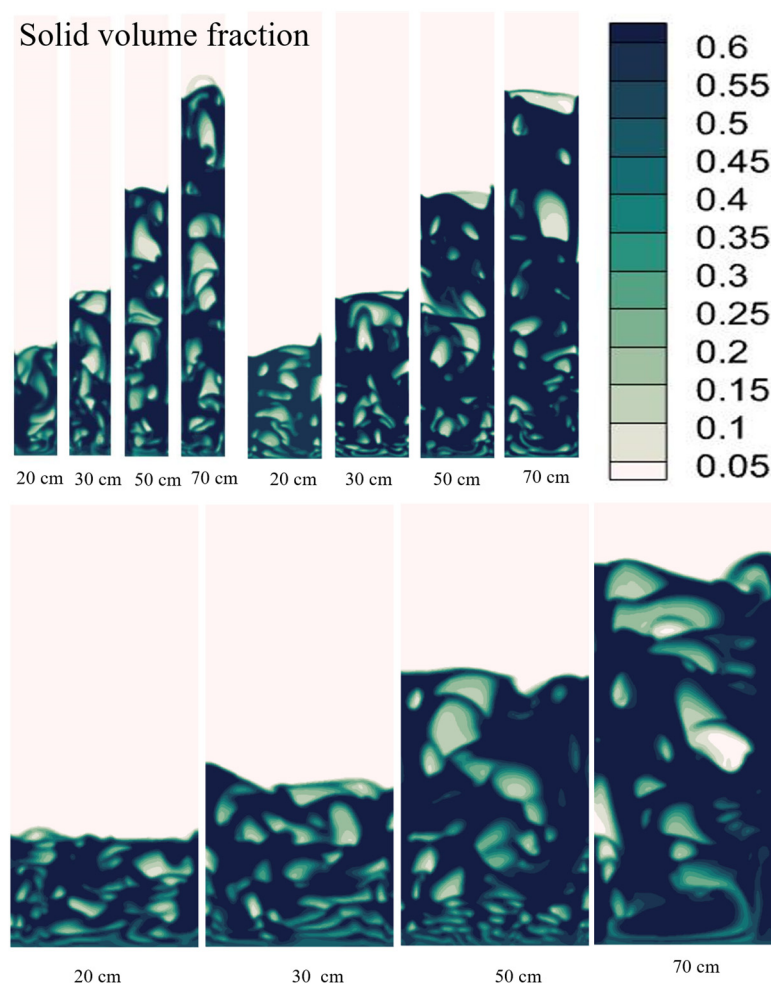


Figure 5. Solid volume fraction changes under various conditions ($t = 5$ s).

3.3. Two-Phase Flow Velocity

Understanding the flow structure of fluidized beds is crucial for the design of fluidized bed separators. The flow pattern of particles within a fluidized bed is a significant factor to consider.

Figures 6 and 7 depict axial velocity changes of solid particles and gas at various bed positions. At $h/H = 0.5$, the gas phase velocity is much greater than the solid phase particle velocity. The velocity at the center is positive, and the velocity at the edge wall is negative. The particles exhibit a circulating flow structure, rising in the middle and falling on both sides within the bed. As the static bed height increases, the cyclic flow structure of gas and solid phases becomes more evident. Furthermore, when comparing small bed diameters, the circulating flow structure at large bed diameters gradually becomes disrupted, resulting in a more uniform and unified flow structure. At $h/H = 1.0$, the gas phase velocity remains greater than the solid phase particle velocity. The impact of the change in static bed height on the axial particle velocity of the gas–solid two-phase flow is not significant. The primary manifestation is a flow structure where gas and solid particles circulate, rising in the center and falling on either side. The impact of the change in bed diameter on the gas–solid two-phase flow is more prominent. As the bed diameter increases, the axial velocity fluctuation of the gas–solid two-phase flow decreases, and the flow structure becomes more uniform in larger bed diameters. Moreover, the influence of bed diameter is greater at $h/H = 1.0$ compared to $h/H = 0.5$. As the fluidized bed size increases, the wall effect weakens, causing smaller velocity fluctuations of particles in larger bed diameters. The graph shows that the gas phase axial velocity has a radial distribution pattern with high values in the middle and

low values on both sides. The low solid content at the center allows gas to pass through easily, leading to this primarily.

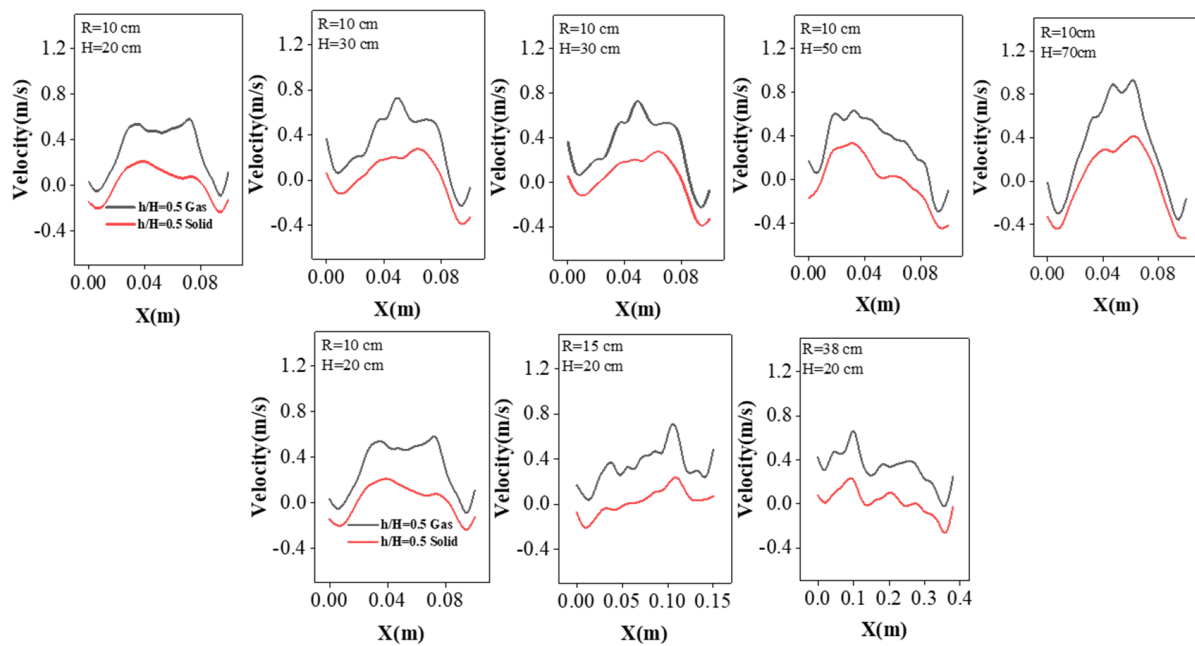


Figure 6. Radial velocity variation of gas and solids ($h/H = 0.5$).

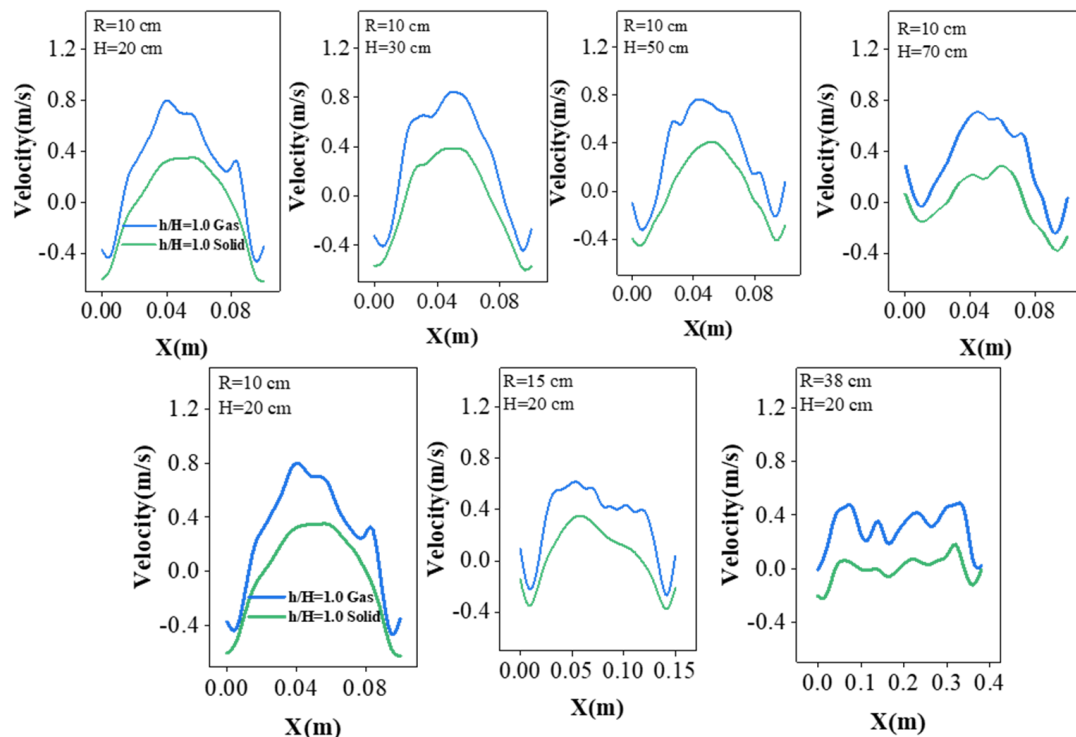


Figure 7. Radial velocity variation of gas and solids ($h/H = 1.0$).

3.4. Radial Density Changes in the Bed

Figure 8a,b show the radial density fluctuations of the bed at various $h/H = 0.5$ positions, with different static bed heights and bed diameters. Based on the graph, it is observed that for a bed diameter of 10 cm, the static bed height does not have a significant impact on the radial density fluctuation of the bed. All variations exhibit similar patterns:

low density at the center of the bed layer and high density near the edge wall. The low density in the central area is due to the edge wall effect, which influences the gas flow through the bed. The central solid content is also low, causing the gas to preferentially pass through the bed along the radial center position. In addition, it is evident from Figure 8b that the alteration in bed diameter has a greater influence on the radial bed density when compared to the variation in static bed height. As the bed diameter increases, the radial density fluctuation of the bed stabilizes. The reason for this is that as the bed diameter increases, it becomes more difficult for the bubble size to match the bed diameter. Small bubbles can pass through the bed not only at the radial center position but also along various pathways away from the edge wall area. As the bed diameter increases, the number of pathways increases and the mutual influence decreases. Therefore, the radial density fluctuation region is stable.

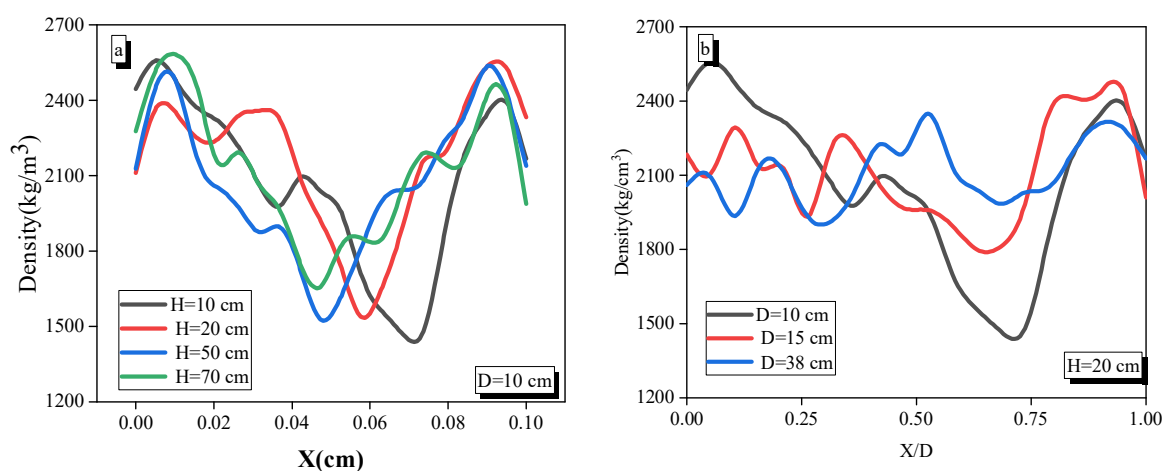


Figure 8. Radial density variation ((a): $D = 10$ cm; (b): $H = 20$ cm).

3.5. Study of Phase Distribution Patterns

Gas–solid fluidization is known for its distinct properties, and the presence of both a bubble phase and emulsion phase in the bed is a prominent fluidization characteristic. The ratio of the bubble phase and emulsified phase composition determines the density and separation performance of the fluidized bed. Therefore, a precise and relevant theory hypothesis is vital for dry separation in fluidized two-phase systems.

Figure 9 shows the solid content fluctuation pattern at the center position of the bed, with a height of $H/2$, when the values of D and H are 10 cm and 70 cm, respectively. We adjusted the bed flow structure to match three distinct instantaneous solid holdups. Orange dots represent particles. Mark point 1: The solid content is small, about 0.1, indicating a larger bubble passing through the test point. Mark point 2: The solid content is relatively high, around 0.6, indicating a large bubble gradually decreasing below the test point. Point 3: The solid content ranges from 0.3 to 0.4, indicating the initial movement of a bubble from the test point. Based on simulation results and existing research, a solid content of 0.3 is selected as the boundary between the bubble phase and the emulsion phase. The solid content of <0.3 represents the bubble phase, while $0.3 < \text{solid content} < 0.6$ indicates the emulsion phase. If any, a solid content > 0.6 corresponds to the agglomerate phase. The plotted quadrants in Figure 10 are shown. Based on the quadrant requirements in the figure, the proportions of the bubble phase, emulsion phase, and agglomerate phase can be determined for various conditions.

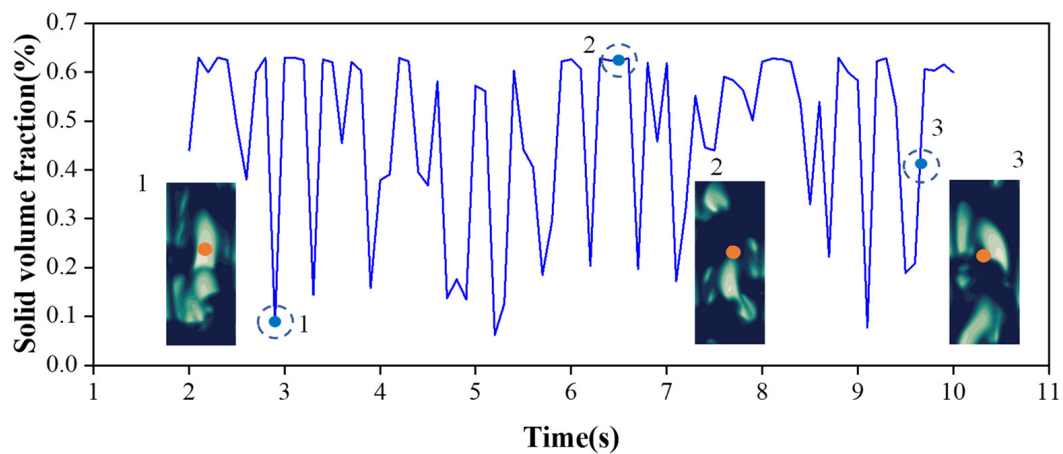


Figure 9. Instantaneous change in solid volume fraction ($D = 10$ cm, $H = 70$ cm).

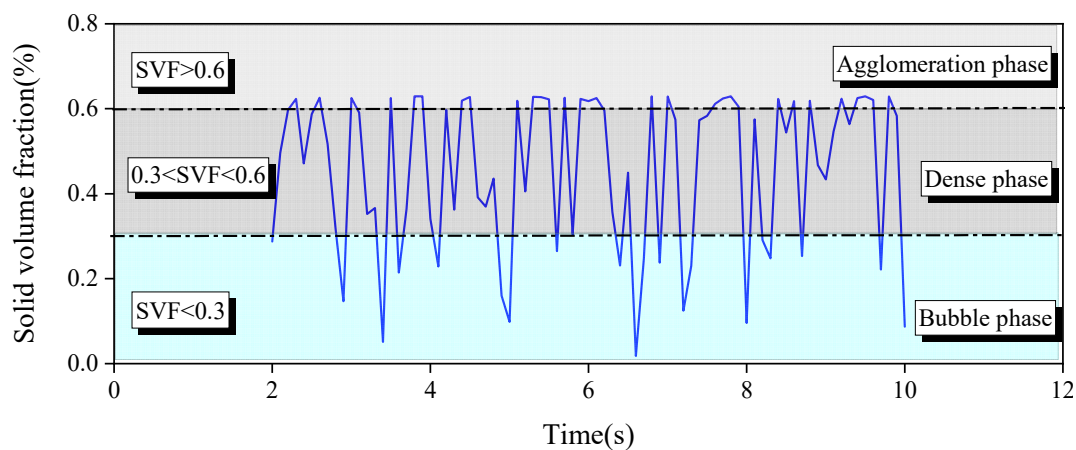


Figure 10. Multiphase partition diagram ($D = 10$ cm, $H = 50$ cm).

Figures 11 and 12 depict the proportional changes of the agglomerate phase, emulsion phase, and bubble phase across various bed diameters and bed heights. Based on the graph, it is evident that with a bed diameter of 10 cm, the proportion of agglomerates increases as the static bed height increases: 25%, 37%, 37%, and 44%. Similarly, the proportions of emulsions decrease to 47%, 38%, 38%, and 35%, and the proportions of bubbles also decrease to 28%, 25%, 25%, and 21%. The ratio of agglomerates increases gradually, while the ratio of bubble phase and emulsion phase decreases gradually. When the bed diameter is 10 cm, as the static bed height increases, the bubble size also increases, and the bubble movement compresses the emulsified phase, leading to agglomerate formation. As the static bed height increases, the disturbance of bubble growth on the bed becomes more apparent. Similarly, as the agglomerate ratio increases, the stability of the bed decreases. With a static bed height of 30 cm, as the bed diameter increases, the proportions of agglomerates are 37%, 0%, and 0%. The proportions of emulsions are 38%, 83%, and 89%. The proportions of bubbles are 25%, 17%, and 11%. This indicates a gradual decrease in the proportions of agglomerates and bubbles and a gradual increase in the proportion of emulsions. In specific static bed height scenarios, when the bubble size is smaller compared to the bed diameter, the bed fluidization becomes more stable. This also leads to smaller bubble disturbance, the gradual disappearance of agglomerates, and an increasing proportion of emulsified phases. Under specific static bed height conditions, it is evident that the greater the diameter of the bed, the higher the likelihood of achieving bed stability.

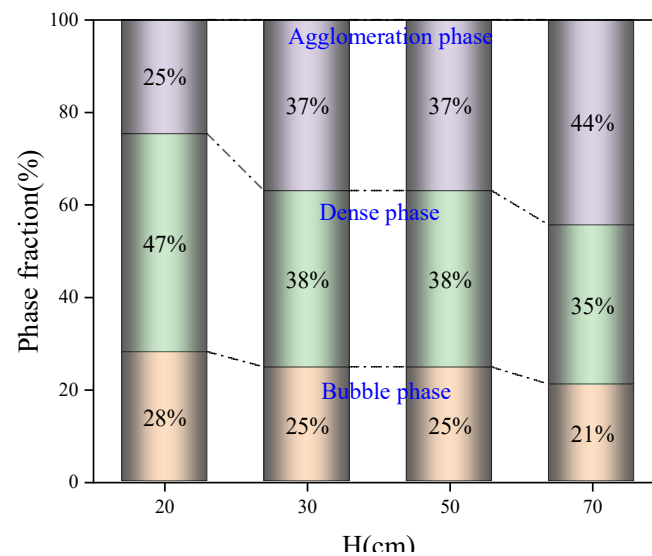


Figure 11. Changes in phase proportions under different conditions ($D = 10$ cm).

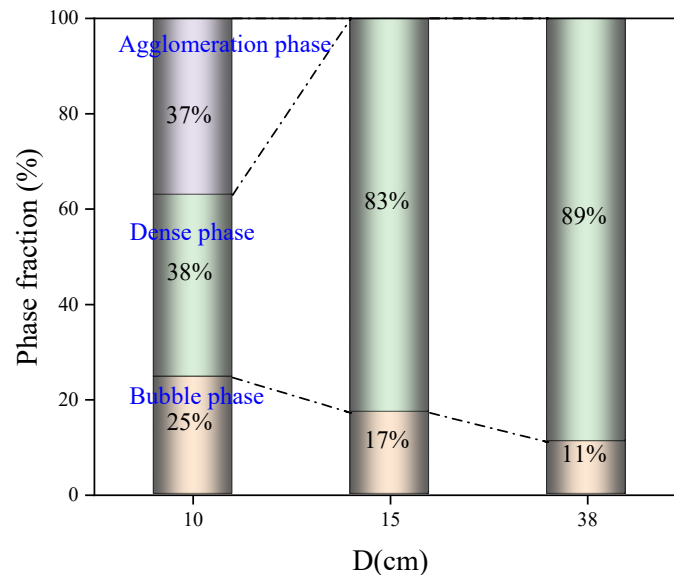


Figure 12. Phase proportion changes under different conditions ($H = 30$ cm).

4. Conclusions

The study analyzed the impact of the continuous enlargement process of a fluidized bed on bed density, gas–solid two-phase flow structure, and proportion changes through simulation. The study yielded innovative findings as follows:

(1) As the bed diameter increases, the amplitude of axial density fluctuations in the bed decreases while keeping the bed height constant. As the bed height increases, the amplitude of axial density fluctuations in the bed also increases while keeping the bed diameter constant.

(2) When the static bed height increases, the cyclic flow structure of gas and solid phases becomes more noticeable. In addition, when comparing small bed diameters to large bed diameters, the circulating flow structure gradually becomes disrupted, resulting in a more uniform and unified flow structure.

(3) As the bed diameter gradually increases, the radial density fluctuation of the bed gradually stabilizes. The impact of altering the bed diameter on radial bed density is more notable when compared to the change in static bed height.

(4) In specific static bed height scenarios, a greater bed diameter promotes enhanced bed stability. As the bed height increases, the proportion of agglomerates also increases. As the bed diameter increases, the bubble disturbance decreases, and the aggregation phase gradually disappears while the proportion of the emulsified phase keeps increasing.

Author Contributions: Conceptualization, Z.G. and H.W.; methodology, Z.G.; software, Z.G. and J.W.; validation, Z.G. and J.W.; formal analysis, H.W.; investigation, H.W.; resources, H.W.; data curation, Z.G. and J.W.; writing—original draft preparation, Z.G. and J.W.; writing—review and editing, Z.G. and J.W.; visualization, H.W.; supervision, C.Z.; project administration, C.Z.; funding acquisition, Z.G. and C.Z. All authors have read and agreed to the published version of the manuscript.

Funding: This research was funded by the Scientific Research Foundation for High-level Talents of Anhui University of Science and Technology (2023yjrc05), National Natural Science Foundation of China (52304279), and National Natural Science Foundation of China (52104276, 52220105008).

Data Availability Statement: The original contributions presented in the study are included in the article, further inquiries can be directed to the corresponding author.

Conflicts of Interest: The authors declare no conflict of interest.

Appendix A

Continuity equation (i = gas phase and solid phase)

$$\frac{\partial}{\partial t}(a_i \rho_i) + \nabla \cdot (a_i \rho_i u_i) = 0 \quad (\text{A1})$$

Momentum equations

For gas phase:

$$\frac{\partial}{\partial t}(a_g \rho_g u_g) + \nabla \cdot (a_g \rho_g u_g u_g) = a_g \nabla \cdot \tau_g + a_g \rho_g g - a_g \nabla p - \beta(u_g - u_s) \quad (\text{A2})$$

For solid phase:

$$\frac{\partial}{\partial t}(a_s \rho_s u_s) + \nabla \cdot (a_s \rho_s u_s u_s) = a_s \nabla \cdot \tau_s + a_s \rho_s g - a_s \nabla p - \beta(u_g - u_s) \quad (\text{A3})$$

Constitutive equations

Stress tensor of gas phase

$$\tau_g = a_g u_g \left[\nabla u_g + (\nabla u_g)^T \right] + a_g \left(\lambda_g - \frac{2}{3} u_g \right) \nabla u_g I \quad (\text{A4})$$

Stress tensor of solid phase

$$\tau_s = a_s u_s \left[\nabla u_s + (\nabla u_s)^T \right] + a_s \left(\lambda_s - \frac{2}{3} u_s \right) \nabla u_s I \quad (\text{A5})$$

Solid pressure

$$P_s = a_s \rho_s \theta_s + 2 \rho_s (1 + e_s) a_s^2 g_0 \theta_s \quad (\text{A6})$$

Radial distribution at contact

$$\frac{1}{g_0} = 1 - \left(\frac{a_s}{a_{s,\max}} \right)^{1/3} \quad (\text{A7})$$

Solid phase bulk viscosity

$$\lambda_s = \frac{4}{3} a_s \rho_s d_s g_0 (1 + e) \left(\frac{\theta_s}{\pi} \right)^{1/2} \quad (\text{A8})$$

Interphase drag coefficient
Gidaspow model:

$$\beta = 150 \frac{a_s^2(1 - a_g)u_g}{a_g^2 d_s^2} + 1.75 \frac{\rho_g a_s |u_s - u_g|}{a_g d_s} \quad a_g \leq 0.8 \quad (\text{A9})$$

$$\beta = \frac{3C_d a_g a_s \rho_g |u_g - u_s|}{4d_s} a^{-2.65} \quad a_g > 0.8 \quad (\text{A10})$$

$$C_d = \begin{cases} \frac{24}{\text{Re}} (1 + 0.15 \text{Re}^{0.687}) & \text{Re} \leq 1000 \\ 0.44 & \text{Re} > 1000 \end{cases} \quad (\text{A11})$$

$$\text{Re} = \frac{\rho_g a_g |u_g - u_s| d_s}{u_g} \quad (\text{A12})$$

Turbulence modeling

$$\frac{\partial}{\partial t}(a_q \rho_q k_q) + \nabla \cdot (a_q \rho_q U_q k_q) = \nabla \cdot (a_q \frac{\mu_{t,q}}{\sigma_k} \nabla k_q) + a_q G_{k,q} - a_q \rho_q \varepsilon_q + a_q \rho_q \Pi k_q \quad (\text{A13})$$

$$\frac{\partial}{\partial t}(a_q \rho_q \varepsilon_q) + \nabla \cdot (a_q \rho_q U_q \varepsilon_q) = \nabla \cdot (a_q \frac{\mu_{t,q}}{\sigma_\varepsilon} \nabla \varepsilon_q) + a_q \frac{\varepsilon_q}{k_q} (C_{1\varepsilon} C_{k,q} - C_{2\varepsilon} \rho_q \varepsilon_q) + a_q \rho_q \Pi \varepsilon_q \quad (\text{A14})$$

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