Powder Technology 308 (2017) 378–387

Contents lists available at ScienceDirect

Powder Technology

journal homepage: www.elsevier.com/locate/powtec

Experimental investigation and correlation of the Bodenstein number in horizontal fluidized beds with internal baffles

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A R T I C L E I N F O

Article history:
Received 5 July 2016
Received in revised form 11 November 2016
Accepted 21 November 2016
Available online 01 December 2016

Keywords:
Continuous horizontal fluidized bed
Particle transport behavior
Internal baffles
Correlation of Bodenstein number
Residence time distribution
Particulate system

A B S T R A C T

In many branches of industry, there is a high demand for the formation or the conditioning of particulate products. In many cases the fluidized bed technology is applied due to its robustness and intensity of heat and mass transfer. Furthermore, many producers look for a continuous process for economic reasons instead of a batch process with high labor and dead-time. These continuous processes, however, hold the disadvantage of a particle residence time distribution (RTD) and therefore a distribution of particle properties. To minimize the spread of the RTD vertical baffles can be installed into the system, causing an influence on the particle transport behavior. To investigate this influence several experiments have been conducted, varying relevant parameters, such as superficial gas velocity, particle diameter, the kind and placement as well as the number of internal baffles, the height of the exit weir and the apparatus itself (pilot and lab-scale plant). From the experiments different influences of the varied parameters on the Bodenstein number, as a measure of convective to dispersive transport, are shown. Based on these trends an empirical correlation for the Bodenstein number is developed and compared to the experimentally determined Bodenstein numbers. Furthermore a correlation for apparent bed porosity is proposed, which uses the set exit weir height instead of the unknown bed height to predict the particle holdup mass and thus the particle mean residence time. It is shown, that both correlations fit well with the experimentally determined values. Finally a prediction of the RTD is shown from the correlated Bodenstein number and the mean particle residence time.

1. Introduction

Particle transport in horizontal fluidized beds is a complex topic, which is investigated for several decades. Its relevance increased even more over the last years, since the production of goods or intermediates increased as well. The intense heat and mass transfer within a fluidized bed makes this process increasingly attractive, especially for chemical and food industries. In a horizontal fluidized bed drying processes of particulate products can be realized, allowing a high throughput of product. Furthermore granulation, coating or agglomeration of particles is possible in such a fluidized bed, by spraying a solution, suspension or melt onto the fluidized particulate material.

Continuous horizontal fluidized bed processes, in particular, holding the advantage of low labor and dead-time, are more and more in demand. However, these processes feature the disadvantage of a particle residence time distribution, which causes a distribution in particle properties. This effect is caused due to the superposition of a forward directed convective transport term and a dispersive transport term, which is directed in all directions. This can be a problem in drying processes, for instance, if particles stay for different periods of time within the system. Particles that leave the system after a short period of time would still possess higher moisture content than particles that stay longer. A similar argument holds for the particle size and its distribution in coating processes.

However, uniformity of properties is a main quality characteristic of particulate products, so that a narrow residence time/property distribution is desired. This means a low deviation of the individual particle residence times has to be reached, which equals low particle dispersion in the horizontal fluidized bed.

One way to do so is to increase the fluidized bed length to width ratio, which is a crucial parameter influencing the spread of the residence time distribution [1]. The larger the ratio, the further the particle transport behavior shifts towards the behavior of ideal plug flow and hence towards a narrower residence time distribution [2]. Due to costs and other practical reasons these long ratios are rarely realized, which is why other methods have to be established to decrease particle dispersion.

Satija et al. [3] reported the installation of baffles into a vibro-fluidized bed to reduce particle dispersion. By doubling the number of baffles

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http://dx.doi.org/10.1016/j.powtec.2016.11.025

Nomenclature

Ar	Archimedes number
Bo	Bodenstein number
C	Concentration kg/m^3
d	Diameter m
D	Dispersion coefficient m^2/s
g	Acceleration of gravity m/s^2
H	Height m
k	Richardson-Zaki exponent
L	Length m
m	Mass kg
$\dot{m}$	Mass flow rate kg/s
n	Number
r	Pearson coefficient
Re	Reynold number
t	Time s
u	Velocity m/s
w	Mass fraction kg/kg
W	Width m
x	Pearson variable
y	Pearson variable
z	Axial coordinate m

Greek letters

ε	Bed porosity
ζ	Dimensionless axial coordinate
θ	Dimensionless time
ν	Kinematic viscosity m^2/s
ρ	Density kg/m^3
$\overline{\tau}$	Mean residence time s

Indices

app	Apparent
b	Bubble
baffle	Internal baffle
bed	Bed
elu	Elutriation
exit	Exit
g	Gas
gap	Gap
in	Inlet
m	Maximum
mf	Minimum fluidization
nb	No baffle
o	Overflow configuration
P	Particle
sample	Sample
sep	Separated
tracer	Tracer
u	Underflow configuration
weir	Weir

Abbreviations

CSTR	Continuous stirred-tank reactor
FB	Fluidized bed
RTD	Residence time distribution

at high vibration amplitudes a decrease of more than 50% was obtained. However, at lower amplitudes no such significant effect was observed.

Nilsson [4,5] investigated the transport of sand and apatite particles in a horizontal fluidized bed of 2.1 m length. He found a definite dependency on the bed height, no influence of the tracer injection point or the tracer mass, but strong influences of the particle diameter and especially the superficial gas velocity on the particle transport behavior and the dispersion.

Due to its importance a tool to predict the particle transport behavior in a fluidized bed is needed. Therefore the Bodenstein number Bo, which quantifies the ratio of convective to dispersive transport, is a suitable number to be correlated.

Even though there are some publications dealing with dispersion [3–6] and its correlation in fluidized beds [1,2,5,7], there exists no correlation, that would take internal baffles into account.

Consequently, the present work extends of a previous work by Bachmann et al. [2] with the focus on the prediction of the Bodenstein number when using internal baffles. Two baffle configurations are investigated: One with particle overflow and one where particles can only flow underneath the internal baffles. In order to use the correlated, dimensionless Bodenstein number, the particle mean residence time $\overline{\tau}$ has to be determined as well, to obtain a dimensional residence time distribution, which could be used for further calculations or simulations for instance.

In general, to calculate the mean residence time an equation by Richardson and Zaki [8] is applied. In the present paper another correlation for apparent bed porosity is proposed to calculate the particle holdup mass and consequently the mean residence time. The advantage of this new correlation is the use of the exit weir height to calculate the particle holdup mass and the mean residence time $\overline{\tau}$, instead of the bed height as required by Richardson and Zaki, which is hard to determine.

2. Experimental setup

The setup and material used in the present work remain the same as described in the previous paper [2] for reasons of comparison. Two plants are used, a lab-scale plant (L_{bed}: 0.64 m; W_{bed}: 0.08 m) and a pilot plant (L_{bed}: 1.0 m; W_{bed}: 0.2 m; Glatt GmbH, GF/Procell 20). The pilot plant is equipped with a heater, for possible hot air supply, whereas the lab-scale plant is not, since it is built of acrylic glass. Both are equipped with two rotary valves to control the particle mass flow and to provide a pressure-tight system. The air flowing through a perforated distributor plate (orifice diameter 1.2 mm; open area 6.4%) and the particle bulk causes fluidization. At the outlet, both plants are equipped with an overflow exit weir. Once particles have been transported over this exit weir they leave the system and cannot re-enter the fluidized bed. A scheme of the fluidized beds is shown in Fig. 1.

The main configuration with respect to the previous publication of Bachmann et al. [2] is the installation of internal vertical baffles in both plants. Two baffle configurations were realized: underflow and