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Experimental study and modeling of particle drying in a continuously-operated horizontal fluidized bed

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ABSTRACT

Multistage fluidized beds are frequently used for product drying in industry. One advantage of these fluidized beds is that they can achieve a high throughput, when operated continuously. In this study, γ-Al_2O_3 particles were dried in a pilot-scale horizontal fluidized bed, without considering any comminution effects. For each experiment, the particle moisture content distribution and residence time distribution were determined. To take into account particle back mixing in our experiments, a one-dimensional population balance model that considers particle residence time was introduced into a fluidized bed-drying model. Experimental particle residence time distributions were reproduced using a tank-in-series model. Subsequently, the moisture content distribution was implemented, as a second dimension to the population balance in this model. These two-dimensional simulations were able to describe the experimental data, especially the spread in the residual particle moisture distribution, much more accurately than one-dimensional simulations. Using this novel two-dimensional model, the effects of different operating parameters (process gas temperature, solid feed rate, superficial air velocity) on the particle moisture content distribution were systematically studied.

Introduction

Particle drying plays an important role in various industries, e.g., chemical, pharmaceutical, agricultural, and food production. Compared with traditional drying techniques (oven drying, tray drying), the fluidized bed has wider scope for particle formation processes (agglomeration, granulation, and coating), in which drying has a significant influence on the final product quality. This reflects the intense hydrodynamic effects of gas–solid flow in the fluidized bed, resulting in strong mixing, as well as strong heat and mass transfer between the two phases.

Fluidized bed systems can be divided into batch and continuous types, based on their operation modes. In a batch fluidized bed, individual particles are processed, with an identical operation time. This type of fluidized bed has been widely investigated for different purposes in the literature (e.g., Burgschweiger, Groenewold, Hirschmann, & Tsotsas, 1999; Fyhr & Kemp, 1999; Hede, Bach, & Jensen, 2009; Rieck, Hoffmann, Bück, Peglow, & Tsotsas, 2015; Srinivas & Setty, 2013). In contrast, solids are continuously fed into and discharged from the process chamber in a continuous fluidized bed. This paper focuses on the horizontal fluidized bed, a common type of continuous fluidized bed, which usually has a long narrow cuboid processing chamber, and continuous movement of the solids in a horizontal direction (Fig. 1). One advantage of this type of fluidized bed is that it can achieve a high yield, when operated continuously. The horizontal fluidized bed is also known as a plug-flow fluidized bed by many researchers (Bizmark, Mostoufi, Sotudeh-Gharebagh, & Ehsani, 2010; Khanali, Rafiee, Jafari, & Banisharif, 2012; Wanjari, Thorat, Baker, & Mujumdar, 2006), because the solid flow should be plug flow under ideal conditions. Under operating conditions, however, the convection of forward-directed particles is disturbed by the dispersion of particles (both forward- and backward-directed). Hence, the ideal solid flow pattern cannot be achieved, and a residence time distribution (RTD) of the solids occurs. As for batch fluidized beds, many experimental studies have focused on horizontal fluidized beds over the past few decades. Reay (1978) developed a correlation for the dispersion coefficient of solids in a horizontal fluidized bed using RTD

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Nomenclature

A	Area (m^2)
Ar	Archimedes number
c	Specific heat capacity (J/(kg K))
C	Heat capacity (J/K)
F	Mass fraction of tracer (kg/kg)
h	Specific enthalpy (J/kg)
Δh_v	Specific enthalpy of evaporation at 0 °C (J/kg)
H	Enthalpy (J)
$\dot{H}$	Enthalpy flow rate (J/s)
J	Number of classes in respect to residence time
K	Theoretical tank number
L	Height (m)
m	Exponent
M	Mass (kg)
$\dot{M}$	Mass flow rate (kg/s)
n	Residence time distribution density (s^{-1})
$\tilde{n}$	Residence time and moisture content distribution density (kg^2/(kg s))
N	Particle number
$\dot{Q}$	Heat flow rate (W)
Re	Reynolds number
t	Time (s)
u	Velocity (m/s)
$\dot{\upsilon}$	Normalized single-particle drying rate
X	Moisture content (solid phase) (kg/kg)
$\bar{X}$	Mean moisture content (solid phase) (kg/kg)
Y	Moisture content (gas phase) (kg/kg)
z	Spatial coordinate (gas phase) (m)

Greek symbols

α	Heat transfer coefficient (W/(m^2 K))
β	Mass transfer coefficient (m/s)
η	Normalized particle moisture content
ν	Kinetic viscosity (m^2/s)
$\dot{\nu}$	Normalized particle moisture content
ξ	Dimensionless spatial coordinate (gas phase)
ρ	Density (kg/m^3)
σ	Standard deviation of particle moisture content distribution (kg/kg)
σ^2	Variance of RTD curve (s^2)
τ	Particle residence time (s)
$\bar{\tau}$	Particle mean residence time (s)
ψ	Porosity
ϑ	Temperature (K)

Subscripts

app	Apparatus
b	Bed
c	Cross sectional
cr	Critical
e	Environment
elu	Elutriation
eq	Equilibrium
exp	Experiment
g	Gas
i	Particle class in respect to moisture content
in	Inflow
j	Particle class in respect to residence time
mf	Minimum fluidization
out	Outflow
p	Particle
t	Time
tot	Total
v	Vapor
w	Water
wet	Wet particle
W	Wall

Abbreviations

LL	Lower limit
BM	Benchmark
CSTR	Continuous stirred tank reactor
NMR	Nuclear magnetic resonance
RTD	Residence time distribution
TIS	Tank-in-series
UL	Upper limit

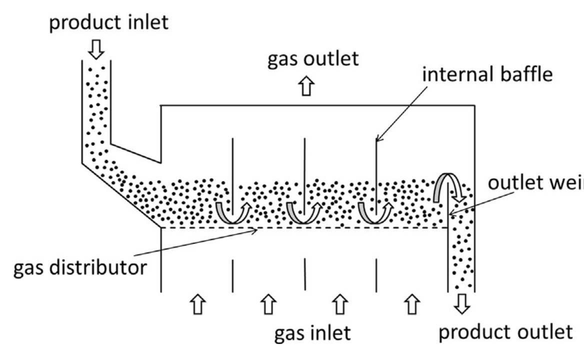

Fig. 1. Schematic diagram of the horizontal fluidized bed used in this study.

experiments with sand and copper. Nilsson and Wimmerstedt (1988) conducted a series of RTD experiments using sand and granulated apatite. They also developed a correlation for the dispersion coefficient. Bachmann, Bück, and Tsotsas (2016, 2017) systematically studied the effects of varying process parameters, such as internal baffle number, superficial air velocity, and outlet weir height, on the RTD of particles. They derived dimensionless correlations using the Bodenstein number (the dimensionless ratio of particle convection to dispersion) under variable particle flow configurations, namely under- and over-flow.

To simulate a horizontal fluidized bed, the two empirical correlations, defined by Reay (1978) and Nilsson and Wimmerstedt (1988), are most often used. As summarized by Daud (2008), conditions between plug flow and ideal back mixing can be described by a series of continuous stirred tank reactors (CSTR) ideal-mixed fluidized beds. However, it is difficult to make an accurate estimation of the number of CSTR fluidized beds required in a simulation. Baker, Khan, Ali, and Damyar (2006) divided the horizontal fluidized bed into several isothermal “drying cells”. Each “cell” was described as a series of CSTRs. The number of CSTRs was determined from the mean residence time and dispersion coefficient, which were calculated using the formula from Nilsson and Wimmerstedt (1988). Wanjari et al. (2006) used a similar approach, but applied the formula developed by Reay (1978) for the dispersion coefficient to calculate the number of ideal-mixed fluidized beds. However, Bachmann et al. (2016) pointed out that the correlations from Reay (1978) and Nilsson and Wimmerstedt (1988) are of questionable validity for other than their own experimental results. Bizmark et al. (2010) calculated the number of CSTRs based on an equivalent Damköhler number (Fogler, 2006), defined as the ratio of the drying rate to the convective rate of moisture removal. Nevertheless, this method has only been validated for a paddy dryer. Another commonly used approach for defining the number of ideal-mixed fluidized beds is based on the geometry of the apparatus, e.g.,