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Short communication

Case study: Optimization of an industrial fluidized bed drying process for large Geldart Type D nylon particles

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Abstract

With an aim to conduct performance rating and optimization of an industrial-scale fluidized bed dryer, a decision tree has been devised to aid selection of the most appropriate mathematical model. The operating fluidization regime is first identified and a theoretical criterion using a cross flow factor is used to decide between using the Class 1 model based on plug flow or the Class 3 model based on the Davidson bubble. In this case-study, this approach is applied on an existing multi-stage dryer for large Geldart Type D nylon particles. The optimization study was carried out using the Class 3 model, which applies the two-phase fluidization theory to determine the transport of moisture between the dense and the bubble phases. An iterative numerical solution has been used to reduce computational time by avoiding the need to solve complex coupled heat and mass balance equations. Changes in bubble size and wall effects along the bed height are taken into account to improve model accuracy. The sensitivity of operating conditions (temperature, weir height, fluidization velocity) and recommendations for optimal operation are presented.

Keywords: Fluidization; Industrial-scale; Drying; Geldart D particles

1. Introduction

Since the first commercial fluidized bed dryer was installed in 1948 to dry dolomite or blast furnace slag [1], hundreds of fluidized bed dryers have been commissioned worldwide to handle a wide range of materials, particularly granular materials which can be readily fluidized such as sand, grains, chemical crystals and fertilizers. Vanecek et al. [2] reported an extensive list of materials in various forms (granular, solutions, suspensions and pastes) that are dried using industrial fluidized beds. Despite its widespread application, the development of efficient and reliable modeling tools for fluidized bed drying has been undermined by a lack of systematic and well-documented data in open literature [3].

For process design and equipment rating of fluidized bed dryers, either the Class 1 plug flow model or Class 3 model based on the Davidson bubble model given by Levelspiel [4] is typically used to model the gas–solids contact pattern.[1] As the plug flow model assumes homogeneous distribution of gas across the cross-section of a bubbling fluidized bed, the actual gas–solids contact can be much poorer than expected. Groenewold and Tsotsas [5] reported that bypassing of gas as bubbles led to lower drying rates in fluidized beds than those predicted using the Class 1 model. This is especially relevant to industrial-scale dryers, where the bed diameter is large and the gas bubbles can grow large enough to reduce the drying rate significantly. Despite such inaccuracies, the plug flow model still remains useful for the process engineer because it is a

[1] As described by Levenspiel [4], historically, three classes of models have been developed to represent the bubbling fluidized bed. The earliest Class 1 model was based on plug flow of gas through the bubbling fluidized bed. It was found that serious bypassing of gas could significantly reduce the gas–solids contact and render the model inaccurate. The Class 2 model divided the fluidized bed into two regions, a dense and a lean solid region, the lean representing the rising bubbles. The model could only fit or summarize experimental data. Described as the best model we have today, the Class 3 model based on the Davidson bubble can predict bubble properties and incorporate the influence of bubble phase on the gas–solids contact inside a fluidized bed.

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conceptually simple model that requires minimal mathematical computation and can be easily found in graduate level textbooks [6–8]. Such inaccuracies can be tolerated in industrial practice by adding a safety margin to over-design the process. In theoretical studies, Hoebink and Rietema [9,10] first reported the use of the Class 3 model to predict drying rates more accurately by accounting for the mass and heat transfer between the gas bubbles and the rest of fluidized bed. Recent work has also focused on coupling two-phase fluidization theory with the relevant drying model and results were verified in laboratory-scale experiments [1,5,11–14]. Such approaches require complex coupled differential equations to be numerically solved for both heat and mass balances, and may be computationally expensive.

The motivation of this paper is to devise a decision tree to aid the selection of the most appropriate mathematical model to rate and to de-bottleneck an existing industrial fluidized bed dryer. A theoretical criterion using a cross-flow factor is developed to select between the two model types according to the fluidization conditions [15]. Comparison of results arising from both model types in predicting the performance of an industrial unit is reported in this paper for the first time. In addition, to render the Class 3 model more amenable and simple to use, a solution scheme has been devised to avoid the need to solve complex coupled differential equations. In modeling the drying of surface water, such equations have been de-coupled into two differential equations, which could be solved relatively quickly via a simple iterative method.

2. Description of industrial process

In polymer production, fluidized bed drying is often the last but a vital operation to assure low moisture content, which is crucial for subsequent polymer processing. The existing industrial-scale dryer is comprised of three sections of fluidized beds in a cascade arrangement as shown in Fig. 1. It is designed to operate as a continuous process with a capacity of two tons solids per hour. Before entering the dryer, nylon is extruded in the form of long strips at approximately 200 °C. These nylon strips are cut into small particles under cooling water before they are charged though a feeder into the dryer, which removes the surface water. The lengths of the three sections are 1 m, 3.5 m and 2.5 m and the width is 1 m. The particle residence time in each section is controlled by varying the weir heights between 0.1 and 0.23 m. A gas distributor made of metal plates perforated with holes is used to keep a low pressure drop of entry fluidizing air. The temperature of entry air into Sections 1 and 2 can be set at 65 °C or 85 °C while that into Section 3 is fixed at 25 °C. The exit air is removed from all three sections via a central exhaust hood. The humidity of entry air is kept low at dew point of 10 °C by chilling it through a cold-trap (heat exchanger). The bed operates slightly above the minimum fluidization velocity of 0.75 m·s^{-1}.

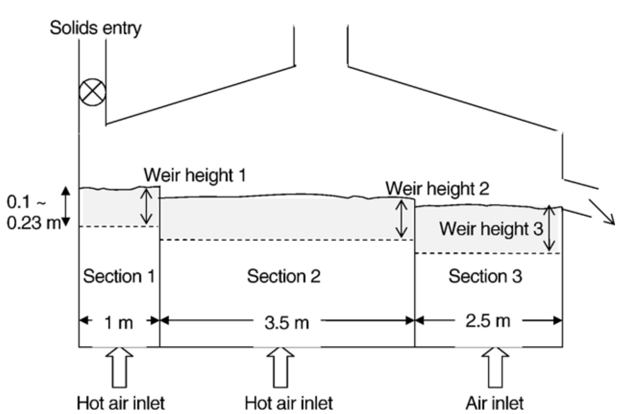

Fig. 1. Schematic layout of industrial fluidized bed dryer.

The nylon particles are elliptical in cross-section and monodisperse. The cross-sectional major and minor diameters are 2.5 mm and 2 mm, and the length is 2.25 mm. Given their relatively large size and high bone-dry density of 1140 kg·m^{-3}, they are classified as belonging to the Geldart D category.

3. Modeling

3.1. Application of the Class 1 model

Assuming plug-flow behavior and neglecting the influence of the bubble phase, the Chilton–Colburn analogy of heat and mass transfer is applied to calculate the transport of surface water from the solid to gas phase. The mass transfer coefficient of water is substituted with a convective heat transfer term h_c/c_{pa} given by the following expressions [8]:

$$h_c = j_H \cdot c_{pa} \cdot \rho_a \cdot u, \text{where} \tag{1}$$

$$j_H = \frac{1.0}{\varepsilon} \cdot \left[\frac{Re_p}{(1-\varepsilon)}\right]^{-0.5} \tag{2}$$

By integrating the change in moisture content inside an infinite small volume over the fluidized bed height, the moisture content at any bed height can be determined by solving Eq. (3) below [16]:

$$\rho_a \cdot u \cdot dY = \frac{h_c \cdot a}{c_{pa}} \cdot [Y_e - Y(H)] \cdot dH \tag{3}$$

By re-arranging Eq. (3), the bed height needed to achieve a given moisture content in the exit air can be expressed in the form of Eq. (4). Eq. (4) is used to calculate the minimum bed height needed to ensure that the moisture content in exit air is maximized, i.e. at adiabatic saturation. As long as the bed heights are higher than the calculated values, the drying rate is determined using the difference in moisture content between the entry air and that at adiabatic saturation.

$$H = \frac{\rho_a \cdot c_{pa} \cdot u}{h_c \cdot a} \cdot \ln\left[\frac{Y_e - Y_i}{Y_e - Y(H)}\right] \tag{4}$$

3.2. Application of the Class 3 model

The overall transfer of surface water between solids and fluidizing gas inside the dense phase is determined using the