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SOLIDS TRANSPORTATION, HEAT AND MASS TRANSFER IN ROTARY DRYERS

By

Yang Song

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ABSTRACT

Rotary dryers are widely used in the process industries. Even though a rotary dryer is in appearance a relatively simple piece of equipment, it is nevertheless very complex as it includes numerous transport phenomena: solids transportation, heat transfer, and mass transfer. To understand the operation of a rotary dryer, these three transport phenomena must be considered simultaneously.

In this thesis, the mechanisms of solids transportation, heat and mass transfer within rotary dryers were first examined. Based on the underlying principles that are used to derive them, solids transportation models in rotary dryers were classified into seven models: empirical model, solid bed motion model, rheological model, cascading solid model, two-stream model, tanks-in-series model, and dynamic model. Heat transfer models in rotary dryers were classified into four categories: empirical model, penetration model, a series of perfectly mixed subsection models, and quasi-3D model. Mass transfer models were also reviewed. Each category of the models was briefly presented and discussed.

Some experimental data obtained in pilot-scale and industrial rotary dryers were used to investigate the influences of moisture content of solids and gas temperature on solids transportation, typically on solids mean residence time

distribution, and on heat and mass transfer to estimate the volumetric heat and mass transfer coefficients.

One pilot-scale rotary dryer with direct contact between the gas and the solids with co-current flow has been designed, constructed and tested in our laboratory. It mainly consists of four parts: an electric fan, an electric heater, a solids feeding system and a rotary cylinder. Numerous experiments were performed to investigate the dynamic characteristics of solids transportation in this pilot-scale rotary dryer. One simple method has been developed to study the influence of various key operation parameters, including solids feed flow rate, gas flow rate, rotation speed and slope of the rotary cylinder, on the residence time of solids within the rotary dryer. Finally a new empirical correlation to predict the residence time was derived from experimental testing.

RÉSUMÉ

Les séchoirs rotatifs sont largement utilisés dans l'industrie de transformation. Même si un séchoir rotatif est en apparence une pièce d"équipement très simple, il est néanmoins très complexe puisqu'il implique plusieurs phénomènes de transfert : transport de solides, transfert de chaleur et de matière. Pour comprendre le fonctionnement d'un séchoir rotatif, ces trois phénomènes doivent être considérés simultanément.

Dans un premier temps, cette thèse examine les mécanismes de transport de solides, et de transfert de chaleur et de matière dans un séchoir rotatif. Sur une base des principes fondamentaux qui ont servi à dériver les modèles de transport de solides dans un four rotatif, ceux-ci peuvent être regroupés sous sept classes : modèle empirique, modèle du mouvement du lit de solides, modèle rhéologique, modèle des cascades du solide, modèle à deux écoulements, modèle des réservoirs en série, et le modèle dynamique. Les modèles de transfert de chaleur dans un séchoir rotatif peuvent être regroupés sous quatre classes: modèle empirique, modèle de pénétration, modèle de réservoirs en série et le modèle quasi-tridimensionnel. Les modèles de transfert de chaleur sont aussi passés en revue. Chaque catégorie de modèles est brièvement examinée et discutée.

Des données expérimentales obtenues lors de tests effectués dans des fours de séchage à l'échelle pilote et industriel ont servi pour étudier l'influence de l'humidité des solides et de la température du gaz sur le transport de solides, et

plus particulièrement pour le calcul du temps de séjour moyen et l'estimation des coefficients volumétriques de transfert de chaleur et de matière.

Un séchoir rotatif, ayant un contact direct co-courant entre le gaz et les solides, a été conçu et construit dans notre laboratoire. Il est principalement constitué des quatre éléments principaux suivants : un ventilateur électrique, une banque d'éléments électriques pour le chauffage, un système d'alimentation des solides et un cylindre rotatif. Un grand nombre d'expériences ont été effectuées pour étudier la dynamique du transport des solides dans ce séchoir rotatif. Une méthode simple a été développée pour étudier l'influence des paramètres importants d'opération, incluant le débit de solides, le débit de gaz, la vitesse de rotation et la pente du cylindre rotatif, sur le temps de séjour dans le séchoir rotatif. Finalement, une nouvelle corrélation pour prédire le temps de séjour a été dérivée à partir de toutes les données expérimentales.

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Foremost, grateful thanks are expressed to my supervisor, Dr. Jules Thibault. His keen insight, fruitful suggestions, ceaseless encouragement, generosity, and great humane understanding were a tremendous source of inspiration and an incredible driving force through the entire course of this research.

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GENERAL NOMENCLATURE

a	Surface area of particles in contact with the gas per unit volume of dryer, m^2/m^3
A	Dryer cross-sectional surface area ($\pi D^2/4$), m^2
A_s	Overall contact surface area between the gas and the solids, m^2
b	Constant
c	Constant
C_p	Specific heat capacity, J/kg K
d	Constant
d_P	Solids particle diameter, m
D	Dryer diameter, m
D_{AB}	Diffusivity, m^2/s
D_p	Weighted average particle size of feed, μm
e	Constant
f	fraction of solid fill [Tscheng and Watkinson, 1979], -
F	Solids feed rate per dryer cross section, kg of dry material/h m^2
F_i	Input solids flow rate, kg of dry material/min
F_o	Discharge or output solids flow rate, kg of dry material/min
G	Gas rate through dryer per dryer cross section, kg/h m2
h	Maximum height of solids in the dryer, m
H	Holdup, kg
k	Conductance, s^{-1}
K	Constant

$K_m a$ Volumetric mass transfer coefficient, kg H_2O/s m^3 Pa

L Dryer length, m

m Mass of wet solids, kg

M Molecular weight, kg/mole

n Speed of rotation, rpm

N Number of Cholette-Cloutier units, -

N_f Number of flights, -

Nu_p Nusselt number (U d_p/k), -

P Pressure, Pa.

Pr Prandtl number ($c_p\mu$/k), -

Q Heat transfer rate, J/s

Q_L Heat losses, J/s

r Dryer radius, m

R Drying rate, kg H_2O/kg dry-solid h)

R_0 Universal gas constant, 8.314 Pa m^3/mol K

Re_p Reynolds number (G d_p/μ), -

Re_T Rotational Reynolds number [Tscheng and Watkinson, 1979], -

R_W Volumetric evaporation rate, kg/m^3s

s Dryer slope, m/m

Sc Schmidt number ($\mu / D_{AB}\rho$), -

t Time, s

T Temperature, K

U Film heat transfer coefficient, W/m^2 K

Ua Volumetric heat transfer coefficient, W/m^3 K

V Volume of cylinder, m^3

X Solids moisture content, kg H^2O/kg dry-solid

z Axial distance, m

Greek letters

α Active zone portion, -

β Mass flow rate exchange ratio between active and dead zones, -

β_a Corrected mass flow rate fraction entering a dead zone, -

β_d Corrected mass flow rate fraction going out of a dead zone, -

γ Angle of repose of material in the dry condition, degree

θ Angle subtended by the bed at the centre of the dryer, rad

ρ Density, kg/m^3

λ Latent heat of vaporization, J/kg

μ Viscosity, kg/m.s

τ Mean residence time, s

Subscripts

a Active zone

d Dead zone

i Indicates properties associated with the i^{th} cell

g Gas

in inlet

out outlet

s Solids

p Particle

w Water

LM Logarithmic mean

Specific to Chapter 3

F Dry solids mass flow rate, kg dry solids/s

G Gas rate through dryer per dryer cross section, kg dry gas/s

H Gas moisture content, kg H_2O/kg dry gas

k Thermal conductivity, W/m K

Specific to Chapter 4

F Solids feed rate per dryer cross section, kg of dry material/min

G Gas rate through dryer per dryer cross section, kg dry gas/min

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CHAPTER 1

INTRODUCTION

1.1 What is a rotary dryer?

Drying is a very common unit operation in industries that is often considered as a harmonious mixture of art and science. Its process is usually very energy-intensive and has a large commercial impact on the cost of the final product. For these reasons, drying has been a constant challenge for engineers and scientists for more than one century. Despite the availability of newer and more specialized dryers, rotary cylinders are still widely used in the process industries not only for drying but also for calcining, clinkering, mixing, and cooling or heating, where drying can occur simultaneously during these processes. Rotary dryers are very versatile and can handle a wide variety of solids such as fertilizers, pharmaceuticals, mineral concentrates, cement, sugar, soybean meal, corn meal, plastics and many others. They are considered easy and reliable to operate in addition of being flexible. Rotary dryers can handle feed solids with broad particle size distributions or solids whose physical properties change significantly during processing. The uniform product quality is promoted by the relatively long and nearly identical residence time, and local efficient mixing of the solid materials within the dryer.

A rotary dryer usually consists of a cylinder that is rotated about its axis at a constant speed. Solids introduced into the upper end of the cylinder move progressively towards the lower or discharge end. Depending on the type of contact between the drying gas and the solids, a rotary dryer may be classified as direct or indirect, and co-current or counter-current. Co-current operations are more often used than counter-current operations even though the latter mode of operation is more energy efficient. The main drawback for counter-current applications is the higher temperature of the final product, resulting in handling problems and possible deterioration of some heat sensitive products. An array of lifting flights of various shapes is commonly installed inside the shell to allow the processing of a larger holdup, to promote good mixing, and to ensure an intimate contact of the solids with the drying gas. The effect of flight design (number of flights, flight dimension and shape) on dryer performance is very complex. These flight configurations vary from spirals to straight baffles. As the dryer rotates, solids are picked up by the flights, lifted on a certain distance around the cylinder, then fall through the drying gas as a cascading curtain. Most of the heat and mass transfer occurs during the free fall of solids from the flights because of the large gas-solids surface contact area. Flight action is also a major contributor for the transportation of solids through the rotary dryer because of the slope of the dryer. One energy source, normally a combustion furnace burning oil or gas, provides a hot gas flowing through the dryer, usually drawn by an exhaust fan. Fine solids particles carried by the hot flowing gas are to be collected by cyclones, bag filters or wet scrubbers.

1.2 Study of rotary dryers

Given that rotary cylinders are used not only for drying but also for various other processes like heating, cooling, and calcining, a considerable number of researchers have worked on two research aspects: solids transportation, and heat and mass transfer. Different models, from simple empirical correlations, through semi-empirical to sophisticate models, were proposed during the last fifty years.

Solids transportation in rotary dryers is undoubtedly the most important aspect in the design of these dryers as it dictates the flow path of the solids inside the dryer and the time of contact between the solids and the hot drying gas. Most of the research done was focused on the determination of the average residence time and solids holdup in rotary dryers. Friedman and Marshall [1949a] presented the first important empirical model. They expressed the average residence time as a function of operating conditions (air and solids flow rates, speed of rotation), geometry of the dryer (diameter, length, slope), and material characteristics (particle diameter). Two other experimental models that use essentially the same parameters were proposed by Sai et al. [1990] and by Alvarez and Shene [1994b]. The latter has the advantage of being also applicable when the slope of the dryer is null. Theoretical and more complex models, such as solid bed motion model [Lehmberg et al., 1977; Woodle and Munro, 1993], rheological model [Perron and Bui, 1990; 1994], cascading solid model [Schofield and Glikin, 1962; Kelly and O'Donnell, 1968; 1977], two-stream model [Matchett and Baker, 1987; Matchett and Sheikh, 1990], tanks-in-series model [Sai et al., 1990; Duchesne et al., 1996], and dynamic model [Shahhosseini et al., 2000; Douglas et al., 1993] were also proposed.

On the area of heat and mass transfer in rotary dryer, most researchers worked on the empirical models represented by either the volumetric heat transfer coefficient [Miller et al., 1942; Friedman and Marshall, 1949b; McCormick, 1962; Myklestad, 1963a; Sharples et al., 1964; Douglas et al., 1993; Alvarez and Shene, 1994a] or the film heat transfer coefficient [Van Krevelen and Hoftijzer, 1949; McAdams, 1951; Ranz and Marshall, 1952; Kuramae and Tanaka, 1977; Hirosue and Shinohara, 1978; Kunii and Levenspiel, 1977; Tscheng and Watkinson, 1979; Chang, 1994; Alvarez and Shene, 1994a]. Some of these correlations are currently widely used by practitioners. Other more complex models include the penetration model that describes the heat transfer between dryer walls and solids [Wes et al., 1976], a series of perfectly mixed subsection models in which the rotary dryer is discretized into a number of equal perfectly mixed subsections and each section or slice of the rotary dryer is treated as a mini-dryer where the feed is the output stream of the preceding subsection [Barr et al., 1989a, 1989b; Douglas et al., 1993; Wild and Smith, 1993; Cook and Cundy, 1995; Cook et al. 1996; Wang et al. 1993], and the quasi-3D model which consists of a one-dimensional axial plug flow model incorporated in a two-dimensional representation of the bed's transverse plane to predict temperature distribution [Boateng and Barr, 1996]. The mass transfer models were usually derived using heat and mass transfer balances by a set of differential equations for any point along the length of the rotary dryer [Myklestad, 1963b; Sharples et al., 1964; Kisakurek, 1982; Platin, 1982; Douglas et al., 1993; Boateng and Barr, 1996; Duchesne et al., 1996; Blumberg and Schlunder, 1996].

Even though rotary dryers have been used in various industrial processes, and the topic of drying characteristics in rotary dryers has been the subject of significant research for many decades, the understanding of the mechanism of rotary drying is still limited. Until now, there is no acceptable model or correlation that would be valid over a wide range of operating conditions for solids transportation or heat and mass transfer. Some research results obtained by various authors are different, and often conflicting. Moreover, most of the studies were performed on small-scale rotary dryers with dry or free-flowing solids. Some recent experimental results with moist solids in industrial and pilot-scale rotary dryers have clearly shown that the moisture content has a large impact on the average residence time. Most existing correlations can underestimate the mean residence time and holdup as much as one order of magnitude because they do not take into account the changing rheological behaviour of the solids as a function of moisture content.

Rotary drying is a very complex process. It includes numerous transport phenomena: solids transportation along the dryer, heat transfer to the solids, and mass transfer within the solid particle and from the solid surface to the flowing gas. Solids transportation in a rotary dryer depends on the operating conditions, the flow path of the solids within the dryer, the characteristics of the solids, and the geometry of the dryer. Heat transfer in a rotary dryer also combines the geometry of the dryer. Heat transfer in a rotary dryer also combines conduction, convection and radiation more or less simultaneously. It is necessary to develop simple laboratory tests to derive new realistic models and correlations that could be used for the design as well as the operation of rotary dryers.

1.3 Objectives

The objectives of this thesis mainly consist of two main segments. The first is the design and construction of a laboratory-scale rotary dryer. The second objective is to experimentally study the dynamic characteristics of solids transportation in this laboratory-scale rotary dryer, and to develop a simple in situ method to rapidly measure the solids average residence time and holdup inside rotary dryers. Using this simple method, it was desired to study the influence of some key operating parameters on the average residence time and to evaluate existing correlations for the prediction of experimental results obtained from all the experimental tests.

1.4 Structure of the thesis

The main body of this thesis contains three papers that are presented in the next three chapters. One paper has been published, one was submitted and the third one will be submitted shortly for publication.

Chapter 2 (first paper) concentrates on solids transportation in rotary dryers. This chapter first examines the mechanism of solids transportation in rotary dryers. Based on a review of the literature relevant to solids transportation in rotary dryers, the solids transportation models were classified into seven models. Each one is briefly presented and discussed. Some experimental results obtained from pilot-scale and industrial rotary dryers were presented to study the characteristics of solids transportation, specially the influence of moisture content of solids on the residence time distribution within rotary dryers.

Chapter 3 (second paper) focuses on heat and mass transfer in rotary dryers. Also, the mechanism of heat and mass transfer in rotary dryers was first examined. With the literature review of heat and mass transfer within rotary dryers, heat transfer models were classified into four categories. Mass transfer models were also reviewed. The influence of the feed solids moisture content and gas temperature on the heat and mass transfer was investigated and the volumetric heat and mass transfer coefficients were estimated.

In Chapter 4 (third paper) the dynamic characteristics of solids transportation in a laboratory-scale rotary dryer is examined. Numerous experiments were performed and a simple in situ method was developed to estimate the mean solids residence time and to study the influence of some key operating variables on the mean residence time.

Finally, a general discussion, conclusions and recommendations are presented in Chapter 5. References are provided at the end of each chapter.

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CHAPTER 2

SOLIDS TRANSPORTATION IN ROTARY DRYERS

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ABSTRACT

Knowledge of solids transportation within rotary dryers is required for dryer modelling and design. In this paper, the mechanism of solids flow in rotary dryers is first examined. Based on the underlying principles of the flow of solids through rotary dryers, solids transportation models were classified into seven models: empirical model, solid bed motion model, rheological model, cascading solid model, two-stream model, tanks-in-series model, and dynamic model. Each category of models is briefly presented and discussed.

Some experimental results obtained within pilot-scale and industrial rotary dryers are then presented. These results clearly show that the values of the solids mean residence time, predicted using most literature correlations, are much smaller than those obtained experimentally. The empirical correlations largely under-predict the mean residence time mainly because they fail to consider the varying characteristics of wet materials along the dryer. Experimental results clearly indicate that the solids residence time distribution (RTD) is strongly influenced by solids moisture content. For some solids, the presence of two distinct humps in the RTD curve is observed. The second hump becomes predominant for solids with higher moisture contents. Further research is needed to develop a new model in order to incorporate those varying characteristics.

(**Key Words:** Solid transportation; Rotary dryers; Residence time distribution (RTD); Modelling)

1. INTRODUCTION

Drying processes are amongst the most common unit operations encountered in a large number of industries. These processes are usually very energy-intensive and, therefore, have a large commercial impact on the cost of the final product. Numerous types of drying equipment are used; each one being better suited for a particular application. When drying large quantities of granular or powdered materials of fairly non-uniform particle size, rotary dryers are often selected. Because of their flexibility, sturdiness and ease of operation, rotary dryers are considered to be the workhorse among drying equipment. They handle small to large feed rates from medium to high temperature operations, especially for solid materials that are not sensitive to temperature and attrition, and that necessitate long retention times. There exist three main categories of rotary dryers that differ in the mode of transferring heat to the solid material for water evaporation: (1) rotary cascade dryers where particulate solids are lifted and cascaded through a co-current or counter-current hot gas stream, (2) rotary louver dryers where the hot gas stream flows through a rolling bed of solids, and (3) rotary steam tube dryers where material is lifted and cascaded through a rotating steam tube bundle. This paper is concerned with the most commonly used rotary dryer: the rotary cascade dryer.

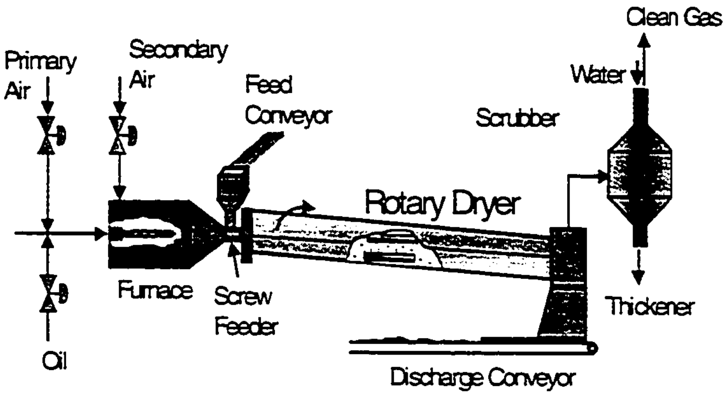

FIGURE 2.1. Schematic Diagram of a Rotary Dryer

Cascade rotary dryers are used in a large number of drying applications: mineral concentrates, salts, sand, agricultural products, lactose, fertiliser, sewage sludge, polymer powder, and numerous other aggregates. The schematic diagram of a typical rotary dryer with co-current gas flow along with other auxiliary equipment is presented in Figure 2.1. Drying is achieved by passing the particulate solids through an inclined rotary cylinder in parallel with a current of hot gas. Lifters or flights attached to the internal surface of the cylinder lift and cascade the solids through the hot gas stream. The lifters contribute to increase the gas-solid contact area and to transport the solids from one end of the dryer to the other due to the angle of inclination of the cylinder. Hot gas flows either co-currently or counter-currently with the solids. Co-current operations are more often used than counter-current operations even though the latter mode of operation is more energy efficient. The main drawback for counter-current applications is the higher temperature of the final product, resulting in handling problems. A combustion furnace, where oil or gas is burned, usually precedes the direct heat dryer to provide a hot drying gas that is drawn through the dryer by an exhaust fan. The drying gas may carry a significant amount of fine solid particles that have to be recovered with cyclones, bag filters or wet scrubbers.

Rotary drying is a very complex process. It includes numerous transport phenomena: solids transportation along the dyer, heat transfer to solids from the flowing gas, and mass transfer within the solid particles and from the solids surface to the flowing gas. Modelling of such a complex process can be very useful for gaining an understanding of the complete drying operation and for process control purposes. To achieve the ultimate objective to devise a comprehensive rotary dryer model, the dynamic representation of the solids transportation must first be obtained since it dictates the amount of time that particles remain in contact with the drying gas. Even though the topic of solids transportation in rotary dryers has been the subject of research for many decades, no satisfactory models or correlations, that would be valid over a wide range of operating conditions, have been proposed to describe the residence time distribution (RTD). In addition, most of the studies were performed on small-scale rotary dryers with dry or free-flowing solids. When the moisture content of solid materials is relatively high, the existing correlations cannot predict the solids mean residence time with sufficient accuracy to be useful for design purposes. There is a clear need to develop new models and correlations to describe the solids transportation of moist solids within a rotary dryer.

This paper describes the research that was undertaken in recent years in our laboratory to develop a new solids transportation model and to better comprehend the influence of solids moisture content on the mean residence time and the shape of the RTD curves in rotary dryers. A brief review of the mechanism of solids flow and solids transportation models in a rotary dryer is first presented.

2. MECHANISM OF SOLIDS FLOW

To better understand the underlying phenomena taking place in rotary dryers, and to determine the optimum operating conditions, it is first necessary to examine the mechanism by which solids flow through a rotary dryer. When a rotary dryer turns on its axis, the solids move from one end of the dryer to the other due mainly to the slope of the drum, and the drag force exerted by the flowing gas. In addition to axial motion, solids are subjected to transverse motion caused by gravitational, frictional, and centrifugal forces. Solid particles, subjected to combined axial and radial motions, move in a rotary dryer assuming very irregular paths.

Axial movement

The axial motion in a rotary dryer is imparted to a solid material by a combination of three or four mechanisms occurring more or less simultaneously, as shown in Figure 2.2. The first and most important mechanism involves *flight or cascade action* [Friedman and Marshall, 1949]. When solids are introduced into a rotary dryer, some portion of solids will continuously be entrained by the rotating flights to the upper half of the cylinder, then fall off the flights back to the lower half of the cylinder. The progressive discharge of the solids from the flights forms a curtain of solid particles perpendicular to the direction of the flowing gas. The point of release of solid particles from the flight is dictated by the angle of repose and friction forces acting on the solid particles. During their fall, solid particles move forward or backward due to the drag force exerted by the flowing gas, and forward due to the inclination of the rotating cylinder. This net forward motion depends on the flight design, solids holdup, slope of cylinder, type of material, and direction and velocity of the drying gas.

The second contribution to the forward motion of solids consists of solids rolling on top of the bed as the cylinder rotates. This action is known as kiln or *rolling action*. Since the rolling action occurs in the direction of the gravitational field and the cylinder is angled to the horizontal, solids move forward. Kelly and O'Donnell [1977] found that, under all operating conditions they investigated, kiln action accounted for more than 40% of the movement of solids through the dryer. The rolling action is favoured by a smaller solids holdup, smaller particles, particles that are more spherical, larger cylinder diameters, fewer numbers of flights, and small angle of repose [Friedman and Marshall, 1949; Kelly and O'Donnell, 1977; Henein et al., 1983a,1983b]. Forces acting on a solid particle in kiln motion are gravitation, friction, and centrifugal forces.

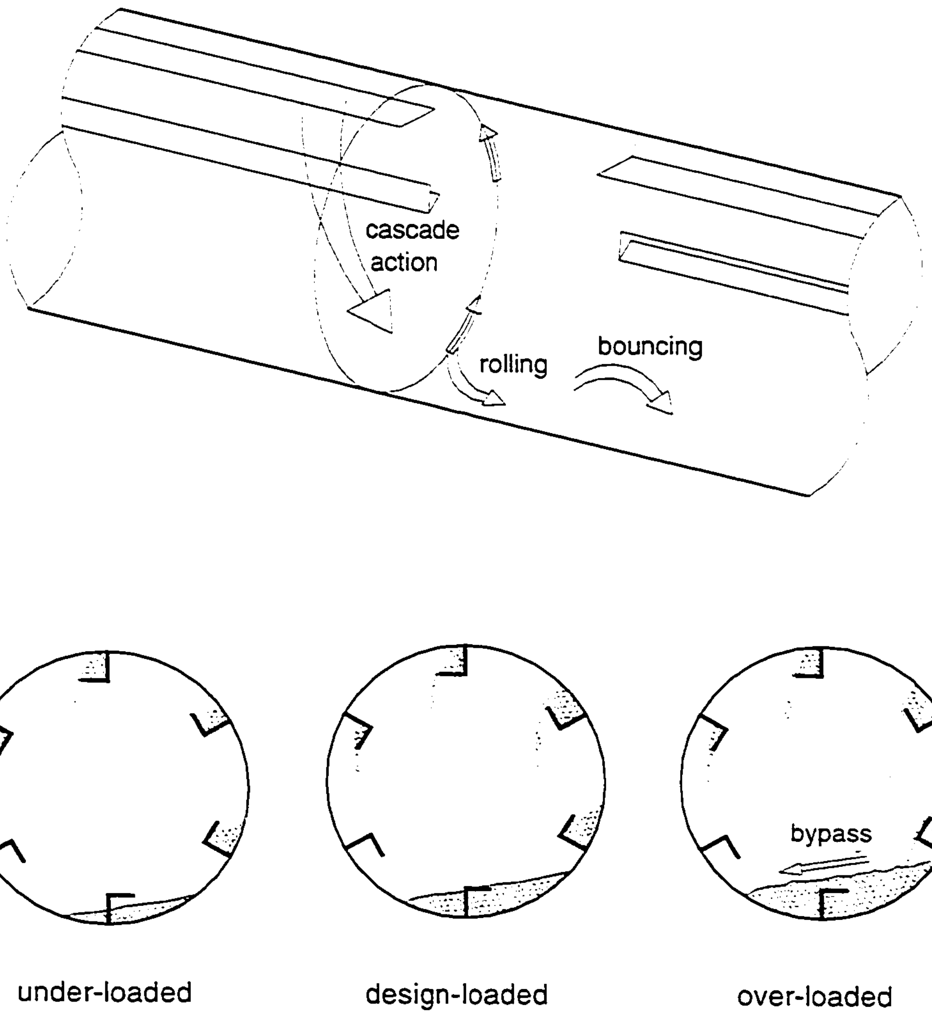

FIGURE 2.2. Schematic Diagram Showing Axial Movement of Solids

When solid particles fall from a flight, it will either strike the drum shell, another flight or the solids bed. Following the impact of falling solid particles on the inclined bed surface, particles may bounce forward along the axis of the dryer and loose their momentum. This third contribution of axial motion is referred to as *bouncing motion* [Kelly and O'Donnell, 1968; Yliniemi, 1999, A]. It may be expected that this mechanism would be more significant for dry or free-flowing solids.

A fourth mechanism to the axial motion of solids may occur when the holdup in the cylinder exceeds the design value. The flights will scrape the lower portion of the bed and, on the average, solids will bypass one or more flights before reaching the lower portion of the bed, where they are entrained by a passing flight and subsequently returned to the upper portion of the solid bed [Kelly and O'Donnell, 1968]. The bed of solid particles progressively sinks from the upper to the lower portion of the bed. In so doing, solids move forward due to the inclination of the cylinder. This minor contribution could be referred as *sinking motion.*

Radial movement

As the transverse motion of solids in a rotary cylinder strongly affects the heat transfer among solids, it has been studied extensively, and several mathematical models of bed motion have been proposed [Henein et al., 1983a, 1983b; Lehmberget al., 1977; Perron and Bui, 1990]. With the rotary drum turning, solids are submitted to transversal motion. Depending on variables such as rotational speed, kiln diameter, feed rate, bed/wall friction, and solid characteristics, the bed motion may adopt one of the many forms: slipping, slumping, rolling, cascading, cataracting and centrifuging as shown in Figure 2.3 [Henein et al., 1983a, 1983b]. As opposed to the previous section, these patterns of solids bed motion refer mostly to rotary cylinders without flights as traditionally encountered in kilns.

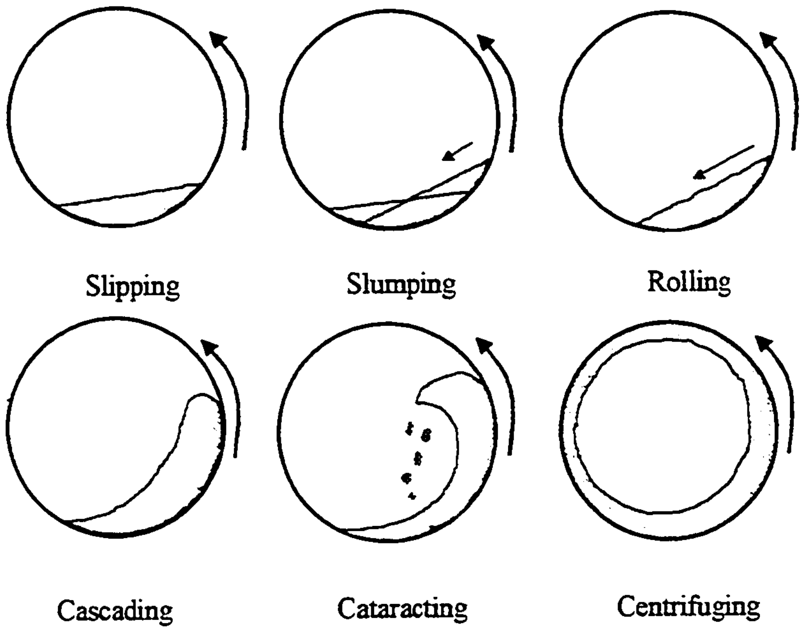

FIGURE 2.3. Schematic Diagram Showing Transverse Bed Motion

At very slow rotational speed, solids in the drum only slip against the inside surface of the drum wall, and very little transverse solid mixing takes place. Increasing the rotational speed, solids movement changes from the slipping to the slumping mode of motion. Slumping motion is similar to an avalanche, where a layer of solids detaches from the upper portion of the bed and slides towards the lower extremity of the bed surface. Slumping occurs in periodical fashion. At a higher rotational speed, the periodic slumping of solids gives way to rolling, characterized by the continuous motion of a layer of solids rolling over the bed surface. With even higher speed of rotation, solids in the upper portion of the bed rise higher up to the wall and begin to fall down in a cascading mode. Higher speeds of rotation propel chunks of materials in the form of a wave (cataracting) or will make the solids adhere to the inside surface of the drum (centrifuging). The latter two modes are not found in practice.

With lifting flights bolted inside the rotating cylinder, most solids progress through the cylinder in a series of cascades, and between cascades, by a sliding or kiln motion between flights down the wall of the cylinder [Mujumdar, 1995].

3. SOLIDS TRANSPORTATION MODEL

Given that rotary drums are used not only for drying but also for various other processes like heating, cooling, and calcining, a considerable number of researchers have worked on solids transportation in rotary cylinders. Most of these researchers have worked on the determination of the average residence time and solids holdup in rotary dryers. Efforts have been devoted to derive acceptable models over a wide range of operating conditions. These models ranged from simple empirical, through semi-empirical, to sophisticate models based on first principles. In this section, an attempt is made to classify the various models based on the underlying principles that were used to derive them. It was not possible to present the contribution of all researchers who have worked on this topic but a judicious selection of references has been provided to cover the most important types of models.

Empirical Models

Empirical models for the estimation of the average residence time and the solids holdup are undoubtedly the most frequently used. The first significant study of the phenomena taking place in a rotary dryer is the one made by Friedman and Marshall [1949]. They performed a large number of experimental tests over a wide range of operating conditions to determine holdup and the average residence time of solids inside a pilot-scale rotary dryer. Variables studied were the slope of the dryer, the speed of rotation, the solids feed and gas mass flow rates, the direction (co-current or counter-current) of the drying gas, and the material characteristics. Friedman and Marshall [1949] used the kiln equation of Sullivan et al. [1927] to derive their empirical correlation for the estimation of the average residence time. The empirical equation of Sullivan et al. [1927], which considers the characteristics of the dryer and the processed solids, is given by the following expression:

$$\tau = \frac{106.2\, L\, \gamma^{0.5}}{\tan^{-1}(s)\, n\, D} \qquad (1)$$

All average residence times in this paper are expressed in seconds so that the constant terms are not identical to those in the original papers. Friedman and Marshall changed this expression by performing a few modifications: (1) the slope of the dryer was represented in units of m/m instead of degrees; (2) a 0.9 power of the rate of rotation was used; and (3) an angle of repose of 40° for the processed solids was assumed. With these three modifications, the expression of Sullivan et al., as reported by Friedman and Marshall, is:

$$\tau = \frac{10.58\,L}{s\,n^{0.9} D} \tag{2}$$

Friedman and Marshall [1949] have modified the previous equation by adding a term to include the effect of the gas and solids mass flow rates, and the particle size diameter on the average residence time:

$$\tau = \frac{10.58\,L}{s\,n^{0.9} D} \pm \left(\frac{608.3}{d_p^{0.5}}\right)\frac{L\,G}{F} \tag{3}$$

The plus sign is for counter-current gas flow whereas the negative sign for the co-current gas flow. This empirical equation was frequently used in the literature and is recommended for the design of rotary dryers [Foust et al., 1960]. Experimental results with no air flow, obtained by Friedman and Marshall, were generally 20 to 40% higher than those calculated with the relationship of Sullivan et al.. Foust et al. [1960] have reported this equation in a nearly identical manner. It appears that Foust et al. have increased the constant in the first term of Equation (3) by 20% to compensate for the error reported by Friedman and Marshall but failed to multiply this first term by 0.6 (60/100) to change the units from hour to minute and from percentage holdup to fraction holdup. As a consequence, the first term of the equation of Friedman and Marshall has in fact been increased by a factor of 98.5% [Renaud et al., 2000].

$$\tau = \frac{21\,L}{s\,n^{0.9} D} \pm \left(\frac{614.2}{d_p^{0.5}}\right)\frac{L\,G}{F} \tag{4}$$

Perry and Green [1984] have reported this equation in the correct manner and the constant in the first term of the equation was increased by 30%, that is the average error that was observed by Friedman and Marshall.

$$\tau = \frac{13.8\,L}{s\,n^{0.9} D} \pm \left(\frac{590.6}{d_p^{0.5}}\right)\frac{L\,G}{F} \tag{5}$$

Another equation for the estimation of the average residence time has been proposed by Sai et al. [1990]. Their model requires knowing a priori the mass holdup in the rotary dryer, which is then used to calculate the average height of solids in the dryer. It is assumed that the solids profile is symmetric, flat, and prevails over the entire length of the cylinder. An estimation of the maximum height of solids has first to be calculated from the holdup in order to use it in the correlation for estimating the mean residence time as given by the following two equations.

$$H = \rho L r^2 (\theta - \sin\theta)/2 \tag{6}$$

$$h = r\left[1 - \cos\left(\frac{\theta}{2}\right)\right] \tag{7}$$

Therefore, knowing the holdup value for particular solids feed rate, inclination and rotational speed of the rotary dryer, the mean residence time can be estimated with:

$$\tau = \frac{78912\,h^{0.24}}{\left[\tan^{-1}(s)\right]^{1.02} n^{0.88} \left(\pi\, r^2 F\right)^{0.072}} \tag{8}$$

This relationship does not consider the influence of the gas flow rate. They found that the mean residence time was nearly independent of feed rate of solids.

More recently, Alvarez and Shene [1994] proposed an empirical expression to estimate the average residence time inside a rotary dryer that can be used even when the slope of the dryer is null.

$$\tau = \frac{7856 d_p^{0.032} \rho^{0.956}}{\pi \ r^2 Fn(18.95s+1)} + \frac{1009 d_p^{-0.065} \rho^{0.002}}{27.22s+1} - \frac{0.108 G^{0.5}}{29.19s+1} \tag{9}$$

Chatterjee et al. [1983a, 1983b] used a dimensional analysis to encapsulate the main operational variables into dimensionless groups to predict the mean residence time of materials in a rotary kiln.

Despite the publication of more sophisticated models, empirical models still remain widely used in correlations for the prediction of the average residence time in rotary dryers. For example, Hatzilyberis and Androutsopoulos [1999a, 1999b] used a very simple equation to predict the average residence time for the flow of lignite particles in bare and baffled rotary dryer. They reported their results only in terms of the rotational speed of the cylinder. In this case, as for many other empirical models, it is difficult to apply these models with confidence to rotary dryers other than those for which the original data were obtained.

Renaud et al. [2000] measured experimentally the mean residence time of mineral concentrate in an industrial rotary dryer and found its value was severely under-predicted by the majority of these empirical correlations.

Model Describing the Motion of the Solid Bed

Numerous studies examined the transverse motion of the solid bed in a rotating drum with no solids flow [Lehmberg et al., 1977; Woodle and Munro, 1993] in order to study the mixing taking place in beds of particles and/or to perform heat and mass transfer studies. Most of these studies were performed without flights so that the cascade motion was absent. Heinen et al.[1983, A, B] have made a thorough study on the various modes of transverse solids motion within a rotary cylinder and have provided a map characterizing each mode with respect to bed depth and rotational speed. Ferron and Singh [1991] used conservation equations to describe planar and bulk regions, coupled by the particulate exchange produced by rotation. Perron and Bui [1994] observed that the motion of a granular bed consists of two modes: an active layer on top of the bed and the bulk of the bed that has an axial velocity significantly lower than the active top layer. They used a rheological model to describe the motion of solids inside a rotating drum without flights (see next section) and to determine the solids axial velocity.

Rheological Model

Perron and Bui [1990] have proposed a very interesting and innovative model. It is based on a study of the motion of solids in the bed to predict the axial velocity of solids as a function of the geometrical and operating parameters of the rotary cylinder as well as the physical properties of solids. This prediction for rotary cylinders without flights is based on dimensional analysis and on the determination of the apparent viscosity of the bed material. The solids are considered to behave as a non-Newtonian and pseudo-plastic fluid. They refined and applied their model to the simulation of the motion of an alumina bed using a computer fluid dynamic code [Perron and Bui, 1990]. Perron and Bui [1994] have extended their model to study the transient responses of the granular bed motion submitted to changes in the speed of rotation and in the angular inclination of the dryer as well as in the solids feed flow rate. The non-linear model was solved numerically with the use of numerous correlations. Their model was tested successfully by Sriram and Sai [1999].

The main difficulty in applying their model is the determination of the apparent viscosity of the bed of solids that is obtained by assuming, or experimentally determining, the velocity profiles associated with the transverse motion of the solid bed. A second difficulty arises in applying this model in the case of drying non-free flowing moist solids where the apparent viscosity and rheological behaviour would drastically change along the length of the dryer. Nevertheless, it is believed that this approach has the merit to consider the physical properties of the solid material and opens the way to the development of a more general equation.

Cascading Solids Model

Some authors performed a force balance, including drag forces exerted by the flowing gas, on the airborne particles released from lifting flights in order to predict the solids average residence time [Schofield and Glikin, 1962; Kelly and O'Donnell, 1968; 1977]. This model assumes spherical particles falling independently of one another. Kamke and Wilson [1986] performed computer simulations using this model and found that cascading solid particles do not in fact behave independently but are influenced by bulk particle flow. Kelly and O'Donnell [1968] observed that solids in a cascade are not subjected to the full force of the gas velocity over the total effective cross section of the cylinder. Indeed, velocities in the space between cascade streams would be higher than in the centre of the cascade. Baker [1992] performed a comprehensive force balance analysis of particles transport through cascading rotary dryers. He found that particles average residence time predictions, considering individual particles, were not accurate compared to experimentally determined values. By treating the curtains of particles falling from individual flights as two-sided flat plates with the drag coefficient calculated accordingly, predictions were much closer to the experimental data. These findings were also corroborated by the comprehensive study of the axial displacement of airborne particles due to drag of the gas stream performed by Sheritt et al. [1994].

Wang et al. [1995] have developed a rigorous mathematical model for the flight discharge rate that considers some of these parameters. Yibin et al. [1999, B] have performed a similar analysis to calculate both the average residence time and the contact heating time of solid particles.

The difficulty in using these models lies in determining the particle flow path. It depends on the local solids holdup, the geometry and number of flights, the kinetic angle of repose, and the average size of particles.

Two-stream Model

Matchett and Baker [1987] proposed a two-stream model that considers both the airborne and dense phases. The airborne phase consists of solid particles falling from the flights and subjected to both gravity and flowing gas drag forces whereas the dense phase consists of solids caught in the flights and on the bottom of the drum. This approach has been considered in other investigations [Saeman and Mitchel, 1954; Glikin, 1978] but not as clearly as performed by Matchett and Baker. They considered cases of rotary dryers being underloaded, design loaded and overloaded. Matchett and Sheik [1990] improved the previous model by taking into account the number and shape of flights in the cylinder. Sheritt et al. [1993] proposed a similar two-stream model and determined the holdup and the axial flow rate of solids in the two phases separately. They claimed that their model is more flexible than previous models.

This two-stream model has been applied to a wide range of literature data to compare predicted and experimental solids average residence times [Matchett and Baker, 1987]. Comparisons were performed for pilot-scale and industrial-scale dryers, with both co-current and counter-current gas flows, and with a wide variety of materials including sands, ion exchange resin, wheat, pumice and fertilizer. The discrepancy between experimental and predicted average residence times was in the range of ±20%. Cao and Langrish [1999] using sorghum grain, performed experiments to validate this model. They observed an average prediction error of -10,4%.

Tanks-in-Series Model

Sai et al. [1990] represented the flow of solids using the dispersion model and a tanks-in-series model, consisting of a large number of equal size back-mix flow tanks. The mean residence time was then correlated with operating parameters and the number of tanks necessary to represent the experimental RTD curves. Duchesne et al. [1996] proposed two tanks-in-series models for the solids transportation in a rotary dryer. The models use the concept of a series of interactive mixers to simulate the flow of solids through the rotary dryer. In the first model, the solids transportation is represented as a fluid passing through a series of perfect mixers, which can be visualised as cylindrical slices of the industrial dryer. This model has only two parameters: the number of perfect mixers N and the conductance k that modulates the flow of solids from one mixer to the next and characterises the ease of the solids to move along the dryer. The conductance is assumed constant for all cells. This model does not account explicitly for the non-ideality of solids flow that can occur in rotary dryers. The ideal flow would be represented by a plug flow system.

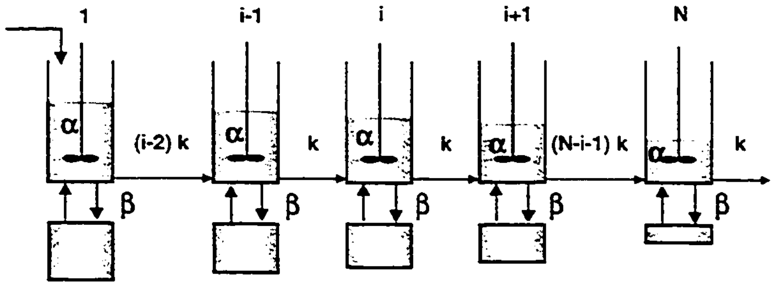

FIGURE 2.4. Modified Cholette-Cloutier solids transportation model

The second model, the modified Cholette-Cloutier model [1959], accounts for the presence of dead zones. The volume occupied by the solids in a cell is divided into two zones: an active zone which contributes to the axial transportation of solids along the dryer and a dead zone where solids are not exchanged with neighbouring cells but only with the corresponding active portion of the cell. This second model requires two additional parameters: α, the fraction of the total volume of solids that is occupied by the active zone, and β, the steady-state ratio of the exchange rate between the active and dead zones to the cell mass feed flow rate. Values of parameters k, α, and β are considered constant for each cell of the rotary dryer. A schematic diagram of the Cholette-Cloutier model is presented in Figure 2.4. For a given slice i of the rotary dryer, the total mass and water balances for both active and dead zones, is given respectively:

$$\frac{d\left(m_{a.i}\right)}{dt}=k\ \left(m_{a.i-1}-m_{a.i}\right)\ \left[1+\beta_{d.i}-\beta_{a.i}\right]-k\ \left(m_{a.i}-m_{a.i+1}\right) \tag{10}$$

$$\frac{d(m_{a.i}w_{a.i})}{dt}=k\ (m_{a.i-1}-m_{a.i})\ \left[w_{a.i-1}+\beta_{d.i}w_{d.i}-\beta_{a.i}w_{a.i}\right]-k\ (m_{a.i}-m_{a.i+1})\ w_{a.i} \tag{11}$$

$$\frac{d(m_{d.i})}{dt}=k(m_{a.i-1}-m_{a.i})[\beta_{a.i}-\beta_{d.i}] \tag{12}$$

$$\frac{d(m_{d.i}w_{d.i})}{dt}=k(m_{a.i-1}-m_{a.i})[\beta_{a.i}w_{a.i}-\beta_{d.i}w_{d.i}] \tag{13}$$

where

$$\beta_{a.i}=\beta+\left(\frac{\frac{m_{a.i}}{m_{a.i}+m_{d.i}}}{\alpha}-1\right) \tag{14}$$

$$\beta_{d.i}=\beta-\left(\frac{\frac{m_{a.i}}{m_{a.i}+m_{d.i}}}{\alpha}-1\right) \tag{15}$$

and

$$\alpha=\frac{m_{a.i}}{m_{a.i}+m_{d.i}} \quad \text{(under steady state)} \tag{16}$$

This model assumes that the active mass fraction, α, is constant and independent for a given solids mass flow rate. When the inlet mass flow rate changes, a new steady state will be achieved with a different rotary dryer loading. As α, the ratio of the mass in the active zone to the total mass in the cell, is constant, the temporary exchange flow rates between active and dead zones were different during a transient period to vary the volume in the dead zones, until the correct value of α is obtained. This rectification is accomplished using equations (14) and (15).

The calibration of the models requires estimating respectively two or four parameters in order to minimise the sum of squares of the errors between the experimental and predicted RTD curves, with the constraint that N can only assume integer values. This model was calibrated by Duchesne et al. [1996] in a 2 m diameter and 15 m long industrial rotary dryer. It was extended by Renaud et al. [2000] by including the variation of parameters of α, β and k as a function of the solids and gas flow rates. This model was also used successfully to represent RTD curves for solids flow in a pilot-scale rotary dryer over a wide range of solids moisture content [Renaud, 2000]. The results showed adequate consistency in both the industrial and pilot-scale dryers. One disadvantage of this semi-empirical model is the necessity to perform tracer tests to calibrate the model. However, it has the advantage to provide both the mean residence time and the RTD curves.

Dynamic model

The majority of models presented in previous sections are steady-state models. Some models have been proposed to study the transient behaviour of the solids within a rotary dryer. Shahhosseini et al. [2000] have divided the length of the rotary dryer into ten segments and the equation of Friedman and Marshall [1949] was applied locally in each segment to calculate the local variation of the holdup. This simple dynamic model was derived for the purpose of model-based process control. Their model was able to follow fairly well the dynamics of key output variables such as the discharge flow rate, and the moisture and temperature of solids. Douglas et al. [1993] performed a similar analysis and also used the equation of Friedman and Marshall [1949]. However, they found that it was not very accurate and had to resort to a multiplicating factor to correct the prediction. Wang et al. [1993] used the identical model without the correction factor. Shene et al. [1996] used a similar approach except that the equation of Alvarez and Shene [1994] was used because it applies for dryers with a null inclination.

The tanks-in-series models that were proposed by Duchesne et al. [1996] are truly intrinsic dynamic models that, once calibrated for a specific dryer, are able to predict very accurately the dynamic flow of solids within rotary dryers. These models represented with a surprising accuracy the flow behavior of solids of a slightly different industrial rotary dryer for which it was calibrated, as evidenced by the results presented in Figure 2.5.

4. RESIDENCE TIME DISTRIBUTION (RTD)

Most correlations and transportation models, described above, were mainly used to estimate the solids mean residence time within rotary dryers. For nearly plug flow of solids, the mean residence time provides adequate information to properly design the rotary dryer. Nearly ideal flow behaviour is encountered for dry or free-flowing solids. For moist solids, the flow behaviour may become highly non ideal and the sole knowledge of the solids mean residence time may be insufficient to properly characterise the rotary dryer. In such situations, the residence time distribution (RTD) is very valuable. To illustrate this point, a series of experiments was performed in both industrial and pilot-scale rotary dryers.

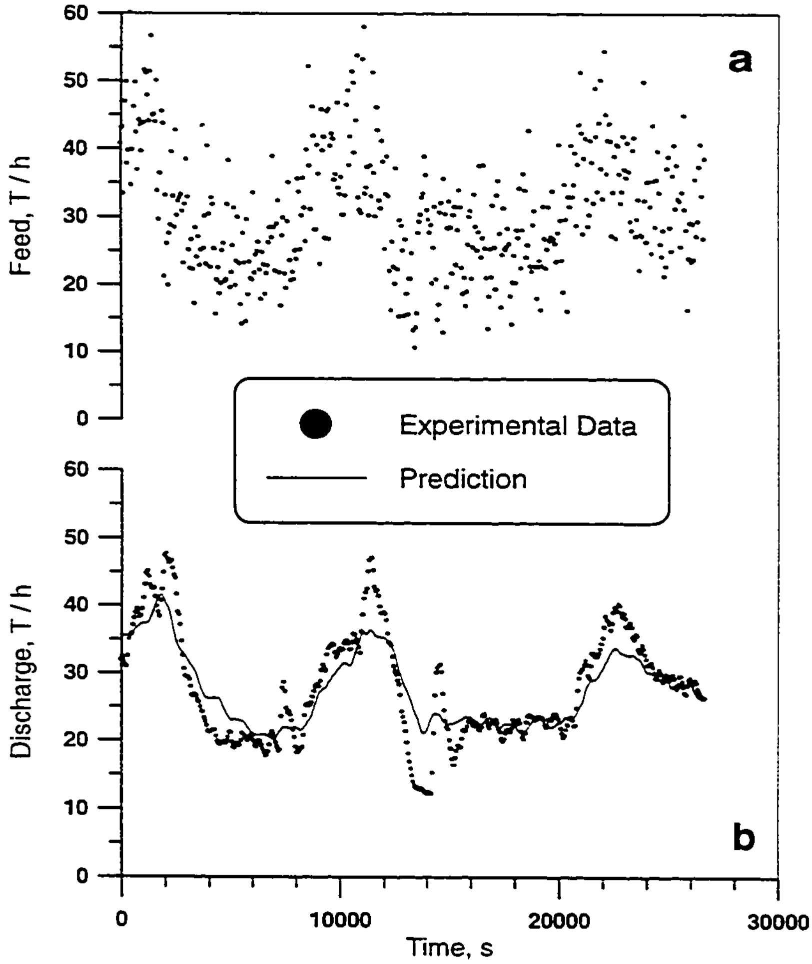

FIGURE 2.5. Dynamic Performances of the Modified Cholette-Cloutier Model:
(a) Measured Concentrate Feed Flow Rate,
(b) Measured and Predicted Discharge Concentrate Flow Rates

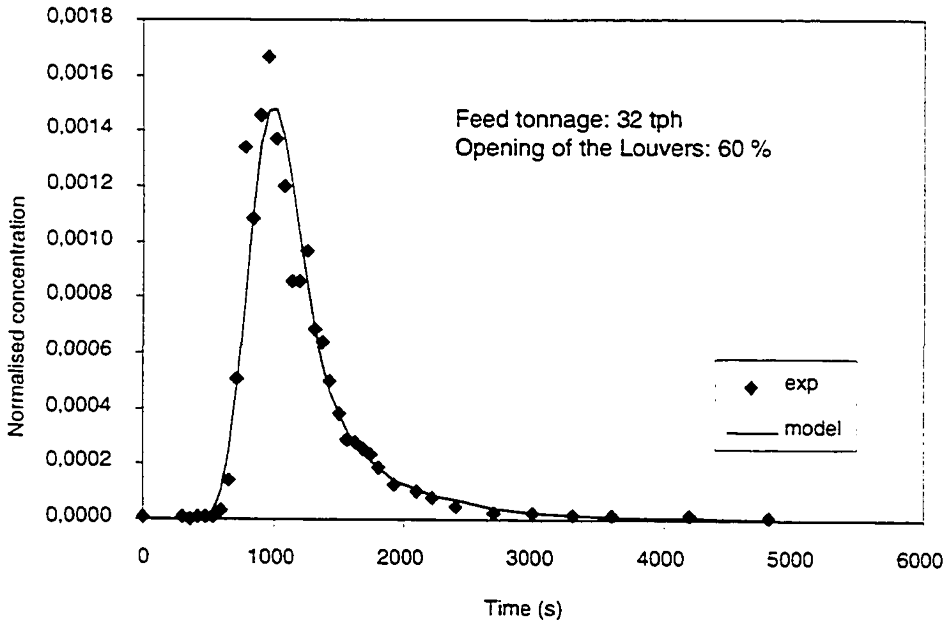

FIGURE 2.6. A Typical Normalized RTD Curve

Figure 2.6 presents a typical normalized RTD curve obtained in an industrial rotary dryer used to reduce the moisture content of a mineral concentrate (zinc sulphide) from roughly 16% to 8% per weight [Renaud et al., 2001]. Lithium chloride was used as the tracer. This curve resembles most RTD curves reported in the literature for dry or free-flowing solids in a pilot-scale rotary dryer. The modified Cholette-Cloutier (solid line) model was able to capture very well the flow behaviour within the rotary dryer including the characteristic extended tail. In passing through the rotary dryer, some portion of the solids is delayed a various number of times by flowing into the dead zones which leads to the extended tail of the RTD curve [Levenspiel, 1972].

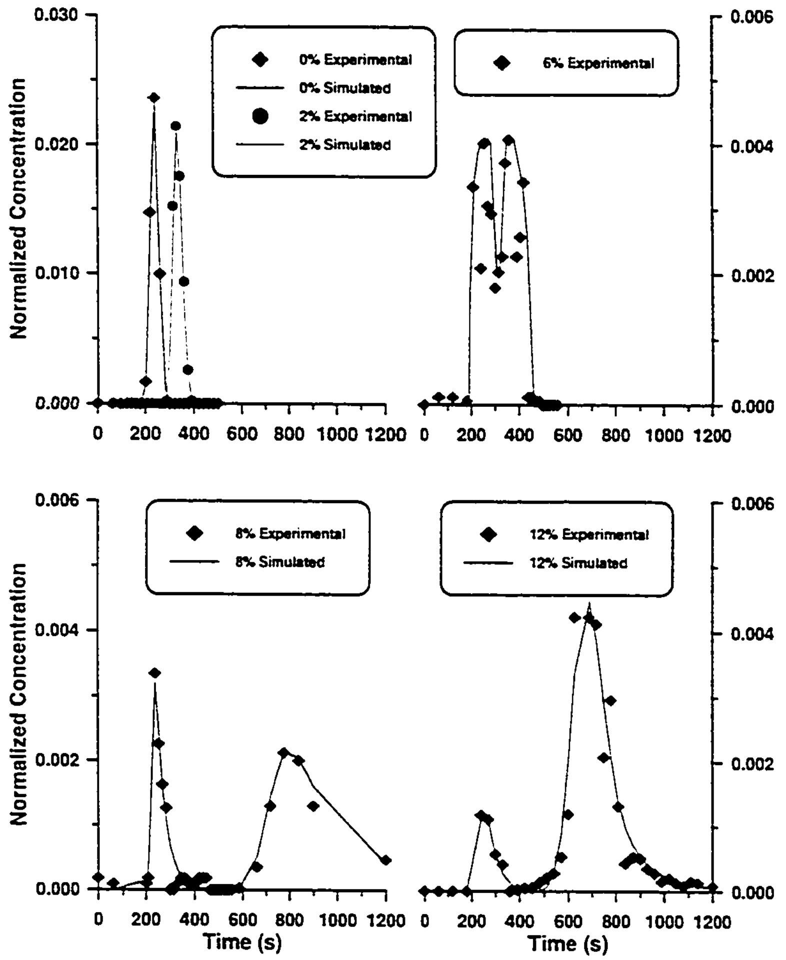

FIGURE 2.7. Residence Time Distribution Curves for 0%, 2%, 6%, 8%, and 12% Solids Moisture Content.

As opposed to the typical behaviour observed with the industrial rotary dryer, tracer tests performed in a pilot-scale rotary dryer (0.31 m in diameter and 3 m long) led to highly non ideal RTD curves [Renaud et al., 2001]. These tests were designed to study the influence of the solids moisture content on both the solids mean residence time and the shape of the RTD curves. Numerous tests were performed with solids feed moisture content varying in the range of 0 to12%, calculated on a wet basis. These tests were performed with air at ambient temperature in order to maintain the moisture content as constant as possible along the dryer to clearly isolate the effect of the moisture content on the mean residence time. Mean residence times calculated from these tracer experiments showed an increase of fourfold when the feed moisture content was increased from 0 to 8%. In the range between 8 to 12% moisture content, no further increase in the mean residence time was observed [Renaud et al., 2001]. Empirical literature correlations failed to estimate adequately the mean residence time.

In addition to the significant variation of the solids mean residence time, the shape of the RTD curves contrasted drastically from typical RTD curves. Figure 2.7 presents RTD curves for 5 different moisture contents. For low feed moisture contents, the shape of the RTD curve approaches the one that would be obtained in a plug flow reactor with relatively low longitudinal dispersion as depicted by the symmetric RTD curve with the relatively low variance. This type of nearly ideal solids flow was observed for 0% and 2% solids feed moisture contents. For higher moisture contents, the shape of the RTD curves departs from ideal flow behaviour. The RTD curves become wider with the presence of an extended tail and almost invariably consist of two humps. The presence of two humps is characteristic of fluid channelling through two parallel paths. For moisture contents of 4% and 6%, almost half of the tracer went out in the first hump. As the moisture content increases, a larger proportion of the tracer is delayed and almost all the tracer exits with the hump having the larger average residence time. The percentages of the tracer exiting with the second hump are 0%, 64%, 80%, and 88% respectively for 0%, 6%, 8%, and 12% moisture contents (see Figure 2.7). The first hump occurred more or less at the average residence time that was observed for dry solids, and the second hump becomes increasingly important and occurs at higher residence time as the moisture content is increased. The hypothesis is that the first hump is associated with solids that cross the dryer by kiln action or rolling mechanism. This action is characterised by the continuous motion of a layer of solids over the bed surface. At high moisture contents, the solid material was sticky and had a tendency to easily attach to the wall of the cylindrical drum. As a result, the action of lifters was significantly hampered. It is hypothesised that the second hump corresponds to solids that attached to the wall, then detached to attach again further along the dryer. The higher the moisture content, the longer it takes to move across the rotary dryer and the larger is the proportion of solids exiting with the second hump. The presence of two humps in a RTD curve in a rotary dryer was previously observed, although not discussed, by Saeman and Mitchel [1954].

These RTD curves would probably be different of those that would be obtained for another solid material in the same pilot-scale dryer or for the same solids in a different rotary dryer. Nevertheless, these results clearly show the major impact that the moisture content has on the solids mean residence time and the importance to include this variable in future correlations. In addition, the transportation behaviour is more a compound effect of the moisture content and the type of solids than moisture content alone. Indeed, Sorghum used by Cao and Langrish [1999] and fish meal used by Alvarez and Shene [1994] were still considered free-flowing with moisture contents of respectively 11.5% and 30% whereas solids used in this investigation ceased to be free-flowing above a moisture content of approximately 4%.

It is important to remind the reader that experiments in the pilot-scale dryer presented above were performed without drying in order to maintain as constant as possible the solids moisture content along the dryer. When the inlet gas temperature increases, drying occurs and the consistency of the solid material changes constantly as it moves along the dryer. As a result, the solids transportation dynamics is enhanced. When the inlet gas temperature is increased to 200°C, the mean residence time decreases by a factor of 2.5 when compared to the value obtained under ambient air temperature. The resulting mean residence time can be viewed as the summation of mean residence times corresponding to local moisture contents along the dryer. When the moist solid material enters the dryer, it has a relatively high mean residence time and a larger holdup. As solids dry, the average time spent inside a given section of the dryer and the local solids holdup decreases. Figure 2.8 presents the RTD curves for the two experiments performed with the initial moisture content of 8% and a temperature of the drying gas of respectively 150°C and 200°C. The corresponding final moisture contents of the two experiments are respectively 2% and 0.6%. In addition to lead to shorter mean residence times and lower final moisture contents, a higher drying gas temperature has a significant effect on the shape of the RTD curves. At a higher drying gas temperature, the RTD curve approaches nearly ideal behaviour akin to free-flowing solids.

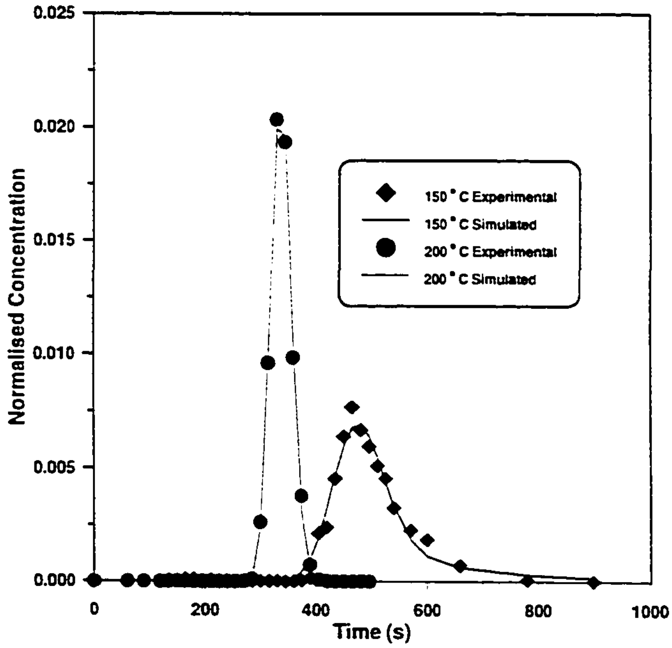

FIGURE 2.8. Residence time distribution curves for 150°C and 200°C drying gas temperature.

The stickiness of moist solids undoubtedly plays an important role in increasing the mean residence time within rotary dryers. Since stickiness is a surface characteristic of solids agglomerates, solids may become free-flowing even if the moisture content is still very high provided that the surface moisture content is relatively low. Therefore, the nature of the solid material, the size of the agglomerates and the distribution of water within the agglomerates, as drying occurs, should be considered in the development of new correlations.

SUMMARY

This paper has attempted to classify the various models that were proposed to describe solids transportation within rotary cylinders. Most models were proposed for the prediction of the solids mean residence time and were mainly developed with dry and free-flowing solids. However, some recent experimental results with moist solids in industrial and pilot-scale rotary dryers have clearly shown that most correlations fail to estimate properly the solids mean residence time because they do not take into account the changing rheological behaviour of the solids as a function of moisture content.

There is a need to develop a new correlation that would include new parameters to account for the characteristics of moist solids in addition to the dryer characteristics and operating conditions. Ideally, it would be nice to rely on simple laboratory tests that would completely characterise, in the form of tractable parameters, solids behaviour as a function of moisture content and of the nature of solids per se. This challenge is now being pursued in our laboratories in view of deriving more realistic models that could be used for the design of rotary dryers.

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CHAPTER 3

HEAT AND MASS TRANSFER IN ROTARY DRYERS

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ABSTRACT

To have a fundamental understanding of the operation of rotary dryers in view of obtaining high quality products and minimizing costs, knowledge of solids transportation, and heat and mass transfer is required. In this paper, the mechanisms of heat and mass transfer in rotary dryers are first examined. Based on the underlying principles, the heat transfer models were classified into four categories: the empirical model, the penetration model, a series of perfectly mixed subsection models, and the quasi-3D model. Mass transfer models were also reviewed. Each category of the models was briefly presented and discussed.

Some experimental data obtained in a pilot-scale rotary dryer were used to investigate the influence of the feed solids moisture content and gas temperature on the heat and mass transfer and to estimate the volumetric heat and mass transfer coefficients. As expected, a higher inlet gas temperature led to a higher drying rate but it was found that the volumetric heat and mass transfer coefficients remained relatively constant. Mass transfer coefficients obtained with the Chilton-Colburn analogy were in close agreement with experimental values.

1. INTRODUCTION

Rotary cylinders are widely used in the process industries for drying, calcining, clinkering, mixing, and cooling or heating. Despite the availability of newer and more specialized dryers, rotary dryers continue to be used in a wide range of applications because they are reliable and easy to operate in addition of being flexible. These dryers can handle feed solids with broad particle size distributions or solids whose physical properties change significantly during processing. The relatively long and uniform residence time of the solid materials within the dryer promotes uniform product quality.

A single pass rotary dryer usually consists of a cylinder that is rotated about its axis at a constant RPM at peripheral speeds from 0.25 to 0.50 m/s. Solids introduced into the upper feed end of the cylinder move progressively towards the lower or discharge end. Depending on the type of contact between the drying gas and the solids, a rotary dryer may be classified as direct or indirect, and co-current or counter-current. An array of lifting flights of various shapes is commonly installed inside the shell to allow processing of a larger holdup, to promote mixing, and to ensure an intimate contact of the solids with the drying gas. The effect of flight arrangement and design (number of flights, flight dimension and shape) on dryer performance is very complex. These flight configurations vary from spirals to straight baffles. As the dryer rotates, solids are picked up by the flights, and lifted on a certain distance around the cylinder before falling through the drying gas as a cascading curtain. Most of the heat and mass transfer occurs during the free fall of solids from the flights because of the large gas-solids surface contact area. Flight action is also a major contributor for the transportation of solids through the rotary dryer.

In spite of simple design, drying in rotary dryers is a very complex process. It includes numerous transport phenomena: solids transportation along the dryer, heat transfer to the solids, and mass transfer within the solid particles and from the solid surface to the flowing gas. Even though rotary dryers have been used for decades in a variety of industrial sectors, general understanding of the influence of the key operating conditions of rotary dryers on the drying performance is still limited. The ability to predict the flow of solids and to estimate the rate of heat and mass transfer within rotary dryers is crucial for modelling and design. The dynamic representation of the solids transportation is paramount to the heat and mass transfer analysis as it dictates the time the solids remain in closed contact with the drying gas. Song and Thibault [2000] examined the mechanism of solids flow pattern and solids transportation in rotary dryers and showed that the residence time distribution is strongly affected by the material moisture content. Since the ultimate objective of drying is to obtain a consistent product moisture content, a clear understanding of the heat and mass transfer mechanisms with respect to the hydrodynamics of the wet material-drying gas along the dryer must be established.

Heat transfer in a rotary dryer occurs as a combination of conduction, convection and radiation. The contribution of each mode to the overall heat transfer rate depends on the operating conditions, the flow path of the solids within the dryer, the characteristics of the solids, and the geometry of the dryer.

The literature review indicates no universal mathematical model or empirical correlation that would allow predicting the heat transfer rate under various operating conditions. It is also important to note that most of the heat and mass transfer correlations for design calculations were obtained on small-scale rotary dryers with dry or free-flowing solids. When the moisture content of solid materials is relatively high, the existing correlations cannot predict the heat and mass transfer with sufficient accuracy as moisture content has a major impact on the solids transportation and thus on transfer processes. To develop a general correlation, it is clear that the correction factor for the wet solids characteristics has to be established.

2. MECHANISM OF HEAT AND MASS TRANSFER

HEAT TRANSFER

The heat transfer process in rotary dryers is highly complex because of conductive, convective and radiative heat transfers that all contribute in different proportions to the overall heat transfer rate from the hot gas to the solids (Figure 3.1).

Conduction

In direct contact rotary dryers, heat from the hot gas is transferred to the surface of solid particles. The resulting surface temperature of a given particle depends on the rate of heat transfer to the particle, the rate of evaporation, and the heat diffusion within the particle. From the point of view of heat conduction involving the solids, three main contributions are present: (1) heat conducted between the particles and the shell/lifters of the rotary cylinder, (2) heat exchanged between contacting particles, and (3) heat conduction inside solid particles. The rate of heat conduction within the particles greatly depends on the thermal properties of solids and its moisture content.

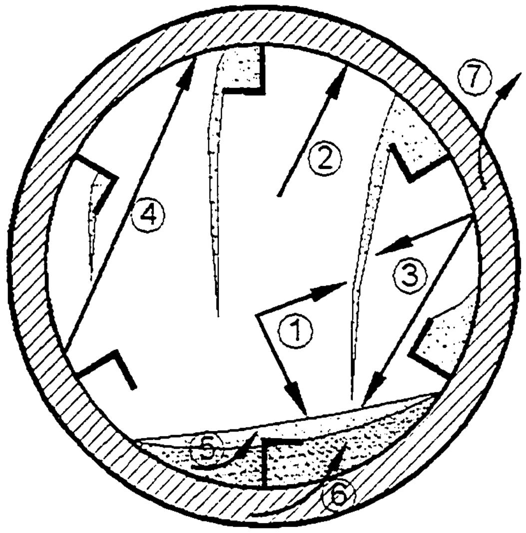

FIGURE 3.1. A cross-section of the rotary dryer showing the various paths for heat transfer.

1: hot gas to exposed bed and cascading solids;
2: hot gas to exposed wall;
3: exposed wall to exposed bed and falling solids;
4: exposed wall to exposed wall;
5: zone to zone within solids bed;
6: dryer wall to solids bed;
7: heat loss to surroundings.

Convection

For most rotary dryers, convection is the predominant heat transfer mode. Upon entering a rotary dryer, the hot drying gas comes in contact with cooler wet solids as well as the cylinder shell. Heat is transferred by convection to the upper surface of the inclined solids bed, and to the curtain of solids falling off the flights. Convective heat transfer rate is difficult to determine since the interfacial area between the hot gas and the solids is unknown. It is a function of the solids-cascade rate, which in turn depends on the size and the number of flights, the holdup and the speed of rotation if flights are used [Saeman and Mitchell, 1954; Matchett and Baker, 1987]. Another non-negligible convective term occurs on the external surface of the rotary cylinder where heat is lost to the surrounding air. Because rotary dryers are rarely insulated, heat losses may be as high as 15% with an extreme of 50% [Friedman and Marshall, 1949]. The rate of heat losses to the environment largely depends on the heat conduction across the dried layer of solids and the solids-to-wall contact area, as typically thermal conductivity of wet solids is higher than the one of dry materials.

Radiation

In rotary dryers where high temperature combustion gases are used and in kilns with a direct flame, the contribution of radiation cannot be neglected. The drying gas contains large amounts of water vapour and carbon dioxide that emit and absorb radiative energy in certain spectral regions. Therefore heat transfer by radiation occurs between the hot gas and the exposed cylinder shell, between the gas and the exposed surface of the bed, between the exposed cylinder shell and the exposed bed, and wall-to-wall exchange.

MASS TRANSFER

The analyses of mass transfer usually represents a greater challenge compared to heat transfer since it is controlled by the latter. Because the latent heat of water evaporation is very large (2500 kJ/kg H_20), the drying rate strongly depends on the ability of the gas to supply heat to the solids. Mass transfer does not only depend on the external heat transfer conditions (temperature, velocity and humidity of the hot air), but also on the internal drying mechanism inside the solids, such as the solids moisture characteristics, thermodynamic equilibrium, and solids sensitivity to temperature [Mujumdar, 1995]. When wet solids are dried, three distinct periods are usually observed: the preheat period, the constant drying-rate period and the falling drying-rate period. Figures 3.2 and 3.3 show respectively a typical variation of the moisture content with time, and the corresponding drying rate curve.

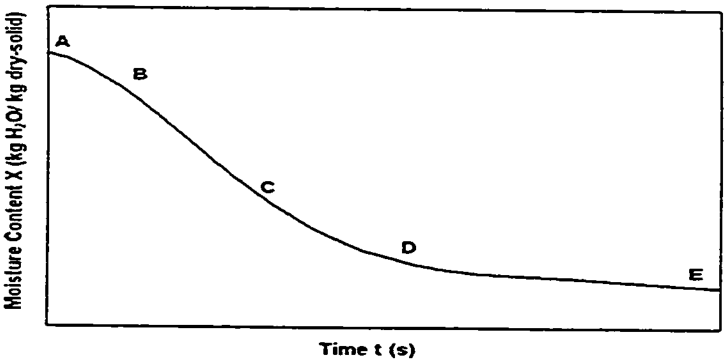

FIGURE 3.2. A typical drying curve showing the moisture content as a function of time (letters are used in the text to describe this curve).

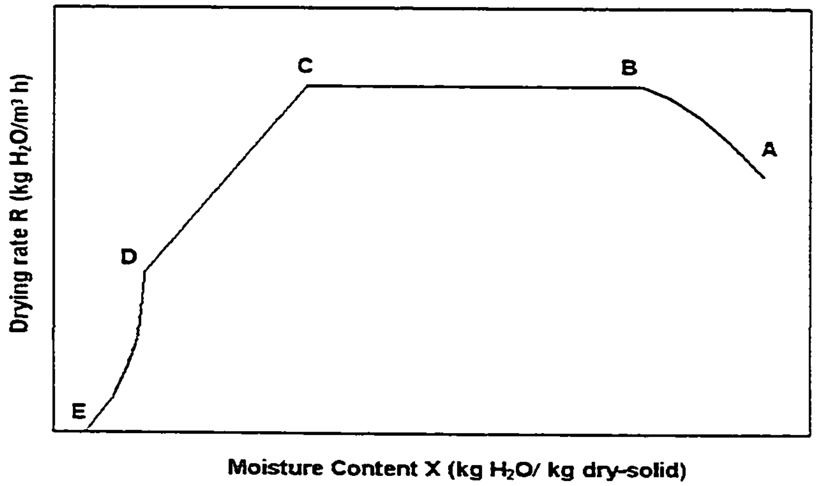

FIGURE 3.3. Plot of the volumetric drying rate as a function of moisture content (letters are used in the text to describe this curve).

After a short unsteady-state adjustment period necessary to increase or decrease the solids surface temperature to its equilibrium value (A-B), a constant rate of drying is usually achieved (B-C). The external heat and mass transfer governs the constant drying-rate period because free water is always available at the surface of evaporation. The surface temperature assumes a constant value such that the rate of heat transfer is exactly equal to the heat necessary to accommodate the rate of water surface evaporation. This constant drying-rate period continues to the critical moisture content where the water can no longer migrate to the evaporation surface at a rate high enough to replenish the solids surface, and the surface becomes progressively unsaturated. Beyond this critical moisture content, the rate of evaporation decreases progressively to zero as the equilibrium moisture content with drying gas is approached C-D-E). This corresponds to the falling drying-rate period, and the drying rate is governed by internal transport limitations.

In a rotary dryer, solids are periodically exposed to high drying rate conditions when solid particles fall through the hot gas, and to low drying rate conditions when being shielded from direct contact with the hot gas by other particles. The three characteristic drying periods are usually observed. The heat transfer Biot number for particulates dried in rotary dryers is usually below 0.2, which implies no temperature gradient throughout the solid particle. Therefore, the rate of evaporation strongly depends on the total solids surface area. The gas-solid interface area is difficult to quantify because most of wet solids is sticky and prone to form large agglomerates. This is the case, for example, of mineral concentrates such as zinc sulphide, or copper sulphide. As solids are dried and tumbled within the rotary dryer, the large agglomerates break, resulting in an increase in the surface area and change in the volumetric heat and mass transfer coefficients along the dryer.

2. HEAT AND MASS TRANSFER MODELS

A considerable number of studies have been done on the development of universally acceptable heat and mass transfer models, and some of these models are widely used by practitioners. These models ranged from the simple empirical, through the semi-empirical, to the sophisticated models. An attempt is made here to classify the various models based on the underlying principles that were used to derive them. It was not possible to present the contribution of all researchers who have worked on this topic but a judicious selection of references has been provided to cover the most important types of models.

EMPIRICAL MODELS

Most previous attempts to analyze heat transfer in rotary dryers have been essentially of an empirical nature, and have been represented by either the volumetric heat transfer coefficient or the film heat transfer coefficient.

Volumetric heat transfer coefficient

The volumetric heat transfer coefficient Ua (W/m^3K) is defined as the rate at which heat is transferred per unit volume of a dryer and per unit temperature difference. The average heat transfer rate (Q) from the gas to the solids is given by the following equation:

$$Q = Ua\ V\ (T_g - T_s)_{LM} \tag{1}$$

where V is the dryer volume and $(T_g - T_s)_{LM}$ is the logarithmic mean temperature difference between the drying gas and the solids at the inlet and outlet of the dryer. This representation eliminates the need to specify the rather nebulous surface area over which heat transfer occurs. The volumetric heat transfer coefficient is usually assumed constant along the dryer.

Numerous correlations that were proposed to estimate the volumetric heat transfer coefficient have been expressed using the following general equation:

$$Ua = K\, f(\varphi)\, G^c\, D^d \qquad (2)$$

where G is the gas mass flow rate (kg/s m^2), D is the internal diameter of the dryer (m); $f(\varphi)$ is a function that includes the influence of numerous hard-to-quantify factors such as properties of the solids, number and geometry of flights, rotational speed and dryer holdup. The parameters K, c and d are empirical constants determined experimentally. A summary of the main correlations that were derived using this empirical equation is presented in Table 3.1.

Table 3.1. Parameters for the different correlations to estimate Ua.

Authors	**K**	**f(φ)**	**c**	**d**	**Note**
Miller et al. [1942]	89	(N_f-1)	0.46	-1	6 flights
	62		0.60	-1	12 flights
Friedman and Marshall [1949]	10	1	0.16	-1	counter-current
	K	1	0.40	-1	co-current
McCormick [1962]	237	1	0.67	-1	
Myklestad [1963a]	423	1	0.80	0	
Sharples et al. [1964]	K	n	0.5	0	
Douglas et al. [1993]	K	$(H100/\rho_s V)^{\beta}$	0.16	0	
Alvarez and Shene [1994]	271	1	0.719	-1	saw dust
	276	1	0.828	-1	sand
	283	1	0.834	-1	soya meal

Most correlations show that the volumetric heat transfer coefficient is inversely proportional to the diameter of the rotary dryer and proportional to the gas flow rate. Parameters K and c vary quite significantly for the different correlations that were derived for a specific set of experiments and under various assumptions. There exists a lot of confusion in the literature in reporting these correlations because of the different systems of units in which they were originally derived and thus erroneous conversion to another system of units. Some authors used the original correlation with different units directly. In this paper, all original equations were converted to SI units.

Despite the general form of Equation 2, various authors have obtained different, and often conflicting, results with regards to the influence of operating variables on the volumetric heat transfer coefficient. Miller et al. [1942] pointed out that the heat transfer coefficient was independent of the rotational speed, the dryer slope, and the solids residence time inside the dryer, whereas McCormick [1962] showed that the geometry of the flight and the speed of rotation should be accounted for in the value of K. The correlation proposed by McCormick gives satisfactory results for steady-state operations, and it is recommended for industrial equipment design in the United States [Treybal, 1980]. Douglas et al. [1993] observed that Equation 2 does not reflect the dynamic behaviour of the heat transfer rate in rotary dryers, and reported that changes of the feed flow rate, along with the rotational speed, the dryer slope and the solids holdup, influence the volumetric heat transfer coefficient. To develop their correlation, Myklestad et al. [1963] assumed that the gas temperature is proportional to the moisture content, and the temperature of solids was constant in the constant drying-rate period but rose linearly with decreasing moisture content in the falling drying-rate period. However in practice these assumptions are not always justified. Furthermore, Alvarez et al. [1994] suggested the following expression for the development of future correlations:

$$Ua = KG^{c} f(N_f) f(\varphi) f(n) \tag{3}$$

where $f(N_f)$ represents the effect of the number of flights, $f(\varphi)$ is the flight geometry factor and f(n) reflects the effect of the rotational speed.

Film heat transfer coefficient

For the volumetric heat transfer coefficient (Ua), the interfacial area between the drying gas and the solids is encapsulated in the overall coefficient. To determine the film heat transfer coefficient (U) from the volumetric heat transfer coefficient, the surface contact area between the gas and the solids must be known or estimated. In a rotary dryer with flights, solids in contact with the drying gas are mainly airborne particles. The film heat transfer approach will be most valid when the number of cascades is large so that the temperature change of a particle during each cascade is relatively small. The heat transfer rate is defined by the following equation:

$$Q = U A_s (T_g - T_s)_{LM} \tag{4}$$

where U is the film heat transfer coefficient (W/m^2K), A_s is the total surface area of all the particles in contact with the drying gas. The use of the film heat transfer coefficient has the advantage to be applicable over a much wider range of operating conditions.

The overall contact surface area (A_S) between the gas and the solids is related to the specific area per unit volume of the dryer (a) by the following equation:

$$A_S = a\,V \tag{5}$$

This surface area greatly depends on the speed of rotation, the number and geometry of flights, the solids holdup, the characteristics of the solids including the particle diameter, the ability to agglomerate, the mechanical strength of agglomerates, and many others.

The estimation of the film heat transfer coefficient is usually generalized with the correlation using dimensionless numbers.

$$Nu_p = \frac{Ud_p}{k} = b + c\,\mathrm{Re}_p^d\,\mathrm{Pr}^e \tag{6}$$

where d_p and k are the diameter and the thermal conductivity of the solids particles, respectively. The constants b, c, d, and e are empirical constants and are obtained by nonlinear regression on experimental data. Table 3.2 presents the empirical constants for a series of correlations proposed in the literature for rotary dryers. It must be noted that the correlations of Alvarez and Shene [1994] were regressed by the authors using their experimental data.

Table 3.2. Values of parameters in the correlations for predicting Nu (Eq. 6).

Authors	b	C	d	e	Note
Van Krevelen and Hoftijzer [Mujumdar, 1995]	0	1.0	0.5	0.33	constant drying-rate period
McAdams [1951]	0	0.33	0.6	0	
Ranz and Marshall [1952]					large particles (d_p>1.4 mm)
Kuramae and Tanaka [1977]	2.0	0.6	1/2	1/3	
Hirosue and Shinohara [1978]					spherical and non-adhesive particles
Kunii and Levenspiel [1977]	0	0.03	1.3	0	small particles
Tscheng and Watkinson [1979]		$0.46Re_T^{0.104}/f^{0.341}$	0.535		from gas to solids
	0	$1.54/Re_T^{0.292}$	0.575	0	from gas to wall
Chang [1994]	0	0.019	1.358	0	
	0	0.065	0.638	0	saw dust
Alvarez and Shene [1994]	0	0.0908	0.394	0	sand
	0	0.0219	0.825	0	soya meal

The equations listed in Table 3.2 are typical correlations that are commonly used for design calculations of heat transfer encountered in heat and mass transfer equipment. However, contrary to heat exchangers, the heat transfer surface area in dryers is poorly defined and the Reynolds number is more difficult to calculate. Yliniemi [1999] found that no universally acceptable model which includes all the relevant design and operating conditions exists, and concluded that available correlations can at least give an order of magnitude of the heat transfer coefficient. Indeed, Hirosue and Shinohara [1978] reported an accuracy of ± 60% when correlating experimental data of many researchers for non-adhesive materials. Correlations are usually valid for drying conditions similar to those for which these correlations were derived. This stresses the need to develop a general equation that will take into account additional operating conditions.

To be able to use the film heat transfer coefficient, some authors have developed equations to estimate the surface area of solids in contact with the gas (A_s). Schofield and Glikin [McAdams, 1951] and Langrish et al. [1988] performed a mass balance for the solids transported by the flights and released in the upper part of the cylinder in order to estimate the mass of solids in contact with the drying gas. These equations include numerous parameters such as the holdup on a flight, the number of flights, the speed of rotation, the average time of the fall of cascading particles, density and diameter of particles, and the length of the dryer.

The definition of the Reynolds number is also not trivial because the gas flow rate seen by solid particles may be different than the average gas flow rate. More importantly, the size of particles will change significantly along the dryer as the moisture content decreases. As noted by Hirosue and Shinohara [1978], correlations may not be applied when adhesive conditions prevail. At higher moisture contents, many solids become sticky and large agglomerates will be exposed to the gas. As the solids dry, the size of agglomerates will progressively decrease and solids will tend to adopt a free flowing behaviour. This change in stickiness of particles influences the heat transfer coefficient, the local solids holdup along the dryer, and the local surface area of solids in contact with the gas.

PENETRATION MODEL

Another heat transfer contribution to the solids is due to the upper section of the cylinder wall heated by the hot drying gas and then coming into contact with the solids. Wes et al. [1976] developed a simple unsteady-state heat transfer penetration model to describe the heat transfer between the wall and solids as they move with the rotating wall. This model assumes that the bulk of the bed is perfectly mixed and that the particles are at the bulk temperature of the bed when they first contact the wall. This model was originally proposed for an externally heated rotary drum. The rate of heat transfer depends on the temperature of the gas and the rotational speed of the cylinder. This heat conduction contribution becomes more important in dryers without flights.

SERIES OF PERFECTLY MIXED SUBSECTION MODELS

Many authors [Barr et al., 1989a, 1989b; Douglas et al., 1993; Wang et al., 1993; Wild and Smith, 1993, A, B; Cook et al., 1995; 1996; Duchesne et al., 1996] have discretized the rotary dryer into a number of equal perfectly mixed subsections. Each section (or slice) of the rotary dryer is considered as a mini-dryer where the feed is the output stream of the preceding subsection. Operating conditions in each subsection are assumed uniform: temperature, gas humidity, solids moisture and drying rate. For each slice of the rotary dryer, complete heat and mass balances are performed in more or less details, expressed as a set of differential equations. Many of these models were derived from experiments with rotary cylinders without flights, but can be adapted for rotary cylinders with flights.

Even though numerous authors used more or less the same approach, each author modelled the rotary dryer using some distinctive features. In the model presented by Wild et al. [1993] for an internally-fired rotary kiln with a premix type burner, the radiative contribution was accounted for using the mean beam length model, and the penetration model was used to include the conductive heat transfer between the wall and the solids. The model proposed by Barr et al. [1989a, 1989b] consists of a finite difference solution for the heat conduction within the wall of the rotating cylinder. The radiative and convective heat transfers within the rotary cylinder served as boundary conditions for the finite difference solution for the wall and accounted for the amount of heat transfer to the solid bed. Cook et al. [1995, 1996] further divided the subsection model into a wall-to-bed sub-model, radiative and convective heat transfer sub-models, and considered water evaporation before the bulk of solids reaches the saturation temperature. Douglas et al. [1993] extended the model to study the dynamic heat and mass transfer using a set of partial differential equations derived from the dynamic energy and mass balances in each section. The dynamic behaviour of a rotary cylinder used to dry sugar was modelled accurately. Wang et al. [1993] developed a non-equilibrium distributed parameter model to represent an industrial sugar dryer. They introduced a correction factor to account for the interaction between the falling particles. They did not assume as in many previous models that the heat transfer between a single falling particle and the drying gas is unaffected by other particles. More recently, Duchesne et al. [1996] proposed a dynamic model which further divided each section of the rotary cylinder into two zones: an active zone which contributes to the axial transportation of solids along the dryer, and a dead zone where solids are not exchanged with neighbouring sections but only with the corresponding active zone of the section. Two additional parameters (K and β) were introduced to represent movement of the solids between two consecutive sections, and between the active and dead zones, respectively. Based on these considerations, this model represents not only the characteristics of heat and mass transfer in both steady and dynamic states, but also accounts for changes in rheological behaviour along the dryer.

QUASI-3D MODEL

Boateng and Barr [1996] proposed a quasi-three-dimensional model to predict heat transfer from the drying gas to the solids bed in a rotary cylinder. This model consists of a one-dimensional axial plug flow model in addition to a two-dimensional representation of the bed's transverse plane. This thermal model is able to predict the temperature distribution within both the solids bed and the cylinder wall at any axial position along the rotary dryer. The solution scheme allows a good predictive model to be obtained without having to account for the complex three-dimensional flow phenomenon.

MASS TRANSFER MODELS

Compared to solids transportation and heat transfer, the literature on mass transfer in rotary dryers is scarce. In addition, many drying experiments were performed by rewetting solids for each experiment with the risk of having a moisture distribution within the solids that is different than the one that would exit in the original solids, and therefore of not adequately representing the industrial applications. However, this potential modification in the solid moisture content should only affect the falling drying-rate period.

The large majority of mathematical models for the mass transfer in rotary dryers have been developed for steady state operations. The models were derived using heat and mass transfer balances for a given point along the length of the rotary dryer. Mass balances are performed on the water content in the solids and in the drying gas. Heat balances are used to account for the change of temperature of the solids and the drying gas. A typical set of differential equations, similar to the one found in Sharples et al. [1964] and Kelly [Mujumdar, 1995], is as follows:

$$
\begin{aligned}
&F\frac{dX}{dz} = -K_m aA(p_s - p_g) = -R_w A \\
&G\frac{dH}{dz} = -F\frac{dX}{dz} = R_w A \\
&F(C_{p,s} + XC_{p,w})\frac{dT_s}{dz} = UaA(T_g - T_s) - R_w \lambda A - \frac{Q_{L,s}}{L} \\
&G(C_{p,g} + HC_{p,v})\frac{dT_g}{dz} = -Ua(T_g - T_s) - R_w C_{p,v}(T_g - T_s) - \frac{Q_{L,g}}{V}
\end{aligned}
\tag{7}
$$

This set of coupled differential equations is normally integrated numerically because many parameters may change significantly with the axial position in the dryer. This may be the case for the volumetric heat and mass transfer coefficients because stickiness of solids may change during drying, thereby resulting in a decrease of particle agglomeration and an increase in available gas-solids surface area. The volumetric drying rate may also change significantly along the length of the dryer as a drying process transits from the constant rate period to the falling one.

Many authors have simplified this set of equations by making various assumptions in view of predicting the solids moisture content as a function of time or dryer length [Myklestad, 1963b; Kisakurek, 1982; Platin et al., 1982; Boateng and Barr, 1996; Blumberg and Schlunder, 1996]. Other authors have considered a dynamic version of this set of differential equations [Douglas et al., 1993; Duchesne et al., 1996].

The convective heat and mass transfer coefficients are also related through the Chilton-Colburn analogy given by the following equation [Holman, 1997]:

$$\frac{Ua}{K_m a} = \rho_g c_p \left(\frac{Sc}{Pr}\right)^{2/3} \frac{R_0 T}{M} \tag{8}$$

This equation allows the estimation of one unknown transfer coefficient using the information for the other transfer coefficient.

WET MATERIALS DRYING

Experiments on heat and mass transfer were conducted in a pilot-scale direct contact rotary dryer, 3 m long and 0.31 m in diameter (Figure 3.4). Six lifters, equally distributed around the circumference, were bolted to the internal surface of the drum with different angles relative to the tangent of the dryer to accommodate the change in rheology of the wet materials. The wet solids were fed directly in the chute of dryer and the solids flow rate was measured by collecting and weighting the solids at the discharge of the dryer over small sampling periods. The average feed mass flow rate for the large number of experiments performed in this investigation was estimated at 67.5 kg/h with a standard deviation of 8 kg/h. The drying gas was a mixture of ambient air and flue gasses from combustion of a propane-butane mixture. The fuel flow rate to the burner was used as the independent manipulated variable to control the inlet gas temperature. The total gas flow rate was controlled by adjusting the louver at the exit end of the suction fan. The gas flow rate was maintained constant at 0.159 m^3/s (STP) during all experiments. The solid material, used in this investigation, was sand that is routinely used in the makeup of cement. The average particle diameter of the dry solid material was 285 μm and its bulk density was 1455 kg/m³.

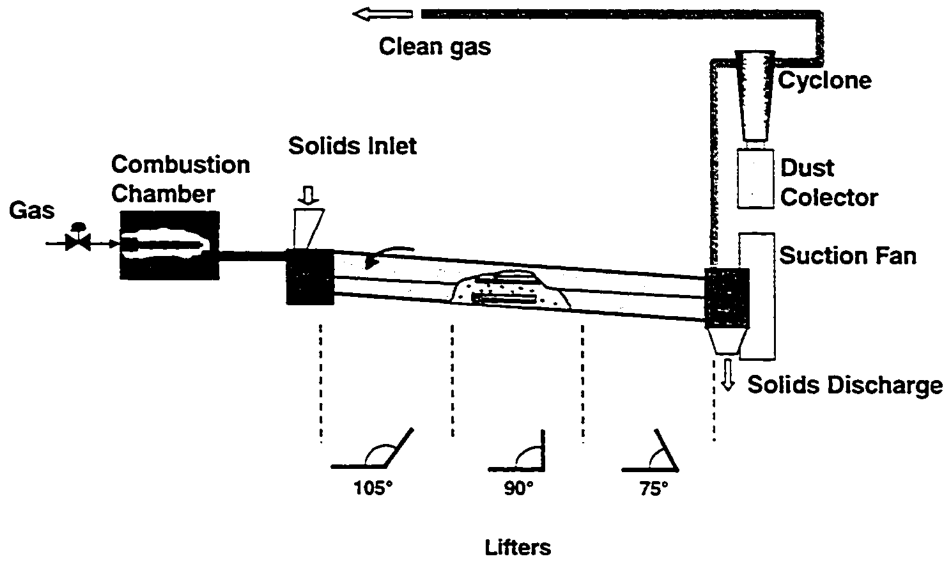

FIGURE 3.4. Schematic diagram of the pilot-scale rotary dryer.

A series of experiments was performed to study the influence of the inlet gas temperature on the drying characteristics of the wet materials and on the volumetric heat transfer coefficient Ua. The solids temperature and moisture profiles and the air temperature profile along the dryer were measured experimentally [Renaud et al., 2001]. The other process variables, such as dryer slope (6.3°), rotation speed (7.0 rpm), solid flow rate and gas flow rate, were kept constant during all experimental runs. Figure 3.5 presents the solids moisture content profile along the rotary dryer as a function of the inlet gas temperature. Using the data of Figure 3.5, plots of the drying rate as a function of solids moisture content were drawn (Figure 3.6). The decrease of the temperature of the drying gas along the dryer with the increase of the solids temperature leads to a decrease in heat transferred to the solids and on the drying rate. Figure 3.6 shows that under current operating conditions, the drying rate is nearly equal to zero at the discharge end of the dryer and the final moisture content only reduces to 0.04, 0.02 and 0.007 kg H_2O/kg dry solids for inlet gas temperatures of 100°C, 150°C and 200°C, respectively.

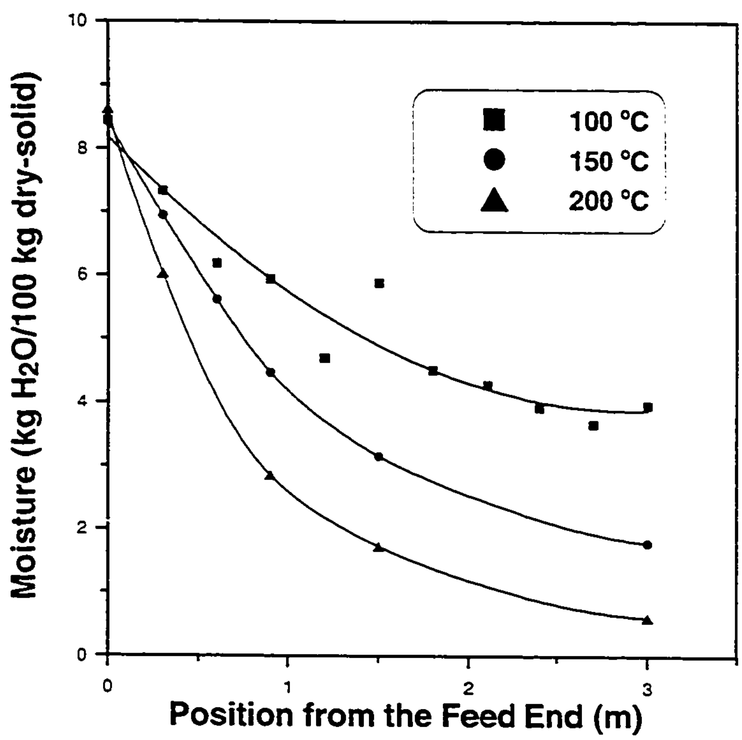

Figure 3.5. Moisture content profile along the rotary dryer for different inlet gas temperatures.

These results clearly show that drying was controlled by the rate at which heat could be provided to the solids. Because both heat and mass transfer coefficients are independent of the inlet gas temperature (see subsequent paragraphs), the higher drying rate can therefore be obtained at higher gas temperatures.

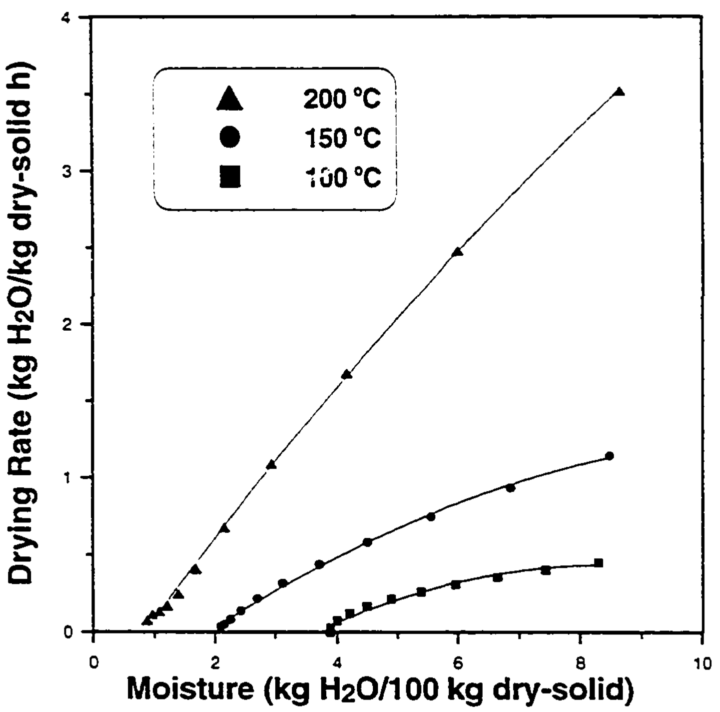

Figure 3.6. Drying rate versus moisture content.

Using the data of Figures 3.5 and 3.6, the volumetric heat and mass transfer coefficients were estimated. For this calculation, it was assumed that all the heat given up by the gas was used for water evaporation. This implies that heat losses to the ambient and the sensible heat for material heating were assumed negligible. The overall volumetric heat transfer coefficient was estimated with the following equation:

$$Ua = \frac{F\lambda(X_{in} - X_{out})}{V(T_g - T_s)_{LM}} \tag{9}$$

where $(T_g - T_s)_{LM}$ is the logarithmic mean temperature difference between the drying gas and the solids at both ends of the dryer. Similarly the overall volumetric mass transfer coefficient is obtained using the logarithmic mean water vapour partial pressure difference between the solids boundary layer and the gas phase:

$$K_m a = \frac{R_w}{(p_s - p_g)_{LM}} \tag{10}$$

It is assumed that the water vapour partial pressure at the gas-solids interface is equal to the saturation partial pressure evaluated using the solids temperature, and the water partial pressure of the inlet gas is equal to zero. The volumetric mass transfer coefficient is also estimated using the Chilton-Colburn analogy (Eq. 8).

Table 3.3. Volumetric heat and transfer coefficients estimated from experimental data.

T_g (in) (°C)	T_g (out) (°C)	T_s (in) (°C)	T_s (out) (°C)	X (in) (kg H_2O/ kg dry-solid)	X (out) (kg H_2O/ kg dry-solid)	Ua (W/m^3.°C)	$K_m a$ (kg H_2O /m^3.s.kPa)	$K_m a$ (kg H_2O/m^3.s.kPa) (C-C Analogy)
100	38	12.3	27	0.0845	0.0382	225.3	0.001838	0.00151
150	54	9.6	36	0.0842	0.021	194.1	0.001986	0.00127
200	67	12	47	0.0862	0.00685	196.5	0.001584	0.00127

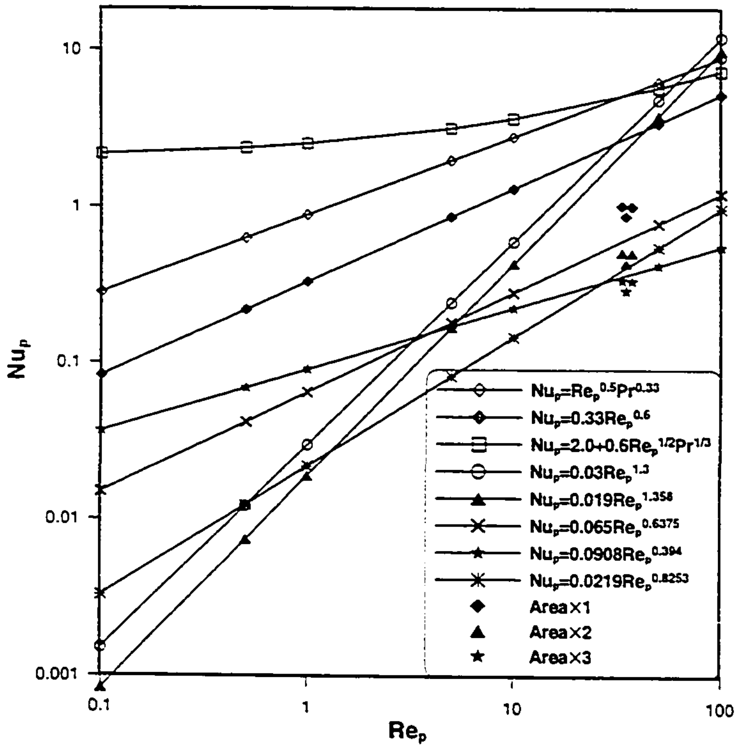

Figure 3.7. Comparison of experimental heat transfer data with existing correlations.

Table 3.3 presents the calculated overall volumetric heat and mass transfer coefficients at three different inlet gas temperatures. These experimental values correspond to typical values reported by others [Friedman and Marshall, 1949; Myklestad, 1963a; Douglas et al., 1993]. The results show that the two transfer coefficients are independent of the drying gas inlet temperature, which is in agreement with the results of Alvarez and Shene [1994]. Figure 3.7 presents a graphical comparison of the experimental Nusselt number corresponding to the results obtained in the pilot-scale rotary dryer with the Nusselt numbers calculated with the correlations listed in Table 3.2. The experimental data were estimated using the area of the solids as if the solids were at rest at the bottom of the dryer as well as twice this area. This type of correlations is quite difficult to use because both the Nusselt and Reynolds numbers are calculated using the average particle diameter of the dry particles whereas in reality the effective particle size should be the average diameter of wet agglomerates moving through the dryer. In addition, an accurate evaluation of the solids-gas surface area is nearly impossible. As the experimental results clearly show that the heat transfer coefficient is very sensitive to the evaluation of the solids surface area in contact with the gas, further research needs to be performed to find ways to properly characterize this surface area. Nevertheless, the experimental results are within the range of few published correlations.

The volumetric mass transfer coefficient obtained using the Chilton-Colburn analogy is close to the one found experimentally. To be valid the Chilton-Colburn analogy requires the physical properties of the fluid stream to be constant [Welty et al., 2001]. This condition cannot be met rigorously in a rotary dryer. The discrepancy is quite small, and is partly due to the widely encountered problems with accurate measurement of the gas humidity at the feed and at the discharge of the dryer. A new rotary dryer is now being constructed with redundant measurements to insure a better estimation of transfer coefficients.

CONCLUSION

This paper reviews the mechanisms of heat and mass transfer within rotary cylinders and attempts to classify the various models published in the open literature. Some preliminary experiments were performed in a pilot-scale rotary dryer to study the heat and mass transfer of wet solids, and to compute the volumetric heat and mass transfer coefficients. Results showed that higher inlet gas temperature had a major impact on the drying rate, whereas the volumetric heat and mass transfer coefficients remained constant.

Most correlations and models for heat and mass transfer in rotary dryers were mainly developed with dry or free-flowing solids. Some recent experimental results with wet solids in industrial and pilot-scale rotary dryers have clearly shown that the residence time is greatly affected by the changing rheological behaviour of the solids as a function of moisture content. The volumetric heat and mass transfer coefficients are also influenced by the solids characteristics. There is a need to develop new correlations that would include additional parameters to account for the characteristics of moist solids in addition to the dryer characteristics and operating conditions.

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CHAPTER 4

DYNAMIC CHARACTERISTICS OF SOLIDS TRANSPORTATION IN ROTARY DRYERS

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ABSTRACT

An original method was proposed for the determination of the mean residence time in a continuous dryer, based on the step-change in the solids feed rate. The method has been validated through experiments performed in a pilot-scale rotary dryer. The effect of the solids flow rate, gas flow rate, dryer rotation speed, and dryer slope was quantified. Several design correlations to predict the residence time in rotary dryers were critically evaluated, and a new, more accurate correlation was derived.

INTRODUCTION

Rotary cylinders have been used successfully for many decades in numerous operations such as drying, calcining, clinkering, mixing, cooling and heating. In drying operations, their widespread use is motivated by their ability to handle a wide variety of particulate solids such as fertilizers, pharmaceuticals, mineral concentrates, cement, sugar, soybean meal, corn meal, plastics and many others. In addition, rotary dryers are known to give uniform product quality because they are characterized by long residence time and relatively good mixing compared to other types of dryers.

The performance of rotary dryers is dictated by three important transport phenomena, namely: solids transportation, and heat and mass transfer. The ability to estimate each of these transport mechanisms is essential for proper design and operation of rotary dryers. It is therefore not surprising that a considerable number of researchers have studied these transport phenomena for many decades. Solids transportation, recently reviewed by Song and Thibault [2000], is undoubtedly the most important one as it dictates the contact time between the solids and the drying gas. The majority of the published literature has considered the determination of the average residence time and solids hold-up. Many models have been proposed, ranging from simple empirical models [Sullivan et al., 1927; Friedman and Marshall, 1949; Foust et al., 1960; Perry and Green, 1984; Sai et al., 1990; Alvarez and Shene, 1994] to comprehensive first principle models [Didriksen, 2002]. Research results have led to an increased understanding of the operation of rotary dryers.

Even though the mechanism of solids transportation has been widely studied, it is still very precarious to estimate the average residence time in many cases. The estimation is sufficiently accurate for dry and free-flowing solids whereas for solids with high moisture content the mean residence time and hold-up can be underestimated by as much as one order of magnitude [Renaud et al., 2000]. In addition, there is a need to develop a simple method to determine rapidly the average residence time and solids hold-up in rotary dryers. An experimental program has been undertaken in our laboratory to achieve these two objectives: (1) to improve existing correlations or propose new correlations to predict the average residence time for various types of solids over an industrially-relevant range of moisture content; (2) to derive a simple *in situ* method to experimentally measure the solids hold-up and the average residence time for industrial rotary dryers. The present paper is mainly concerned with the second objective.

To meet this objective, a simple dynamic experiment based on a step-change of the material feed rate has been developed and validated in a laboratory rotary dryer. This dynamic experiment was then used to determine the influence of the solids flow rate, the gas flow rate, the speed of rotation and the slope of the dryer on the mean residence time and solids hold-up.

MATERIALS AND METHODS

Pilot-Scale Rotary Dryer

A pilot-scale rotary dryer was built in our laboratory to investigate the mechanism of solids transportation. The schematic diagram of the pilot-scale rotary dryer and its auxiliary equipment is presented in Figure 4.1. It is a direct contact dryer with co-current flow between the gas and the solids. The stainless steel rotary cylinder is 3.05 m long and 0.305 m in diameter. It is insulated with aluminium insulating foils to reduce heat losses. The cylinder rests on rubber wheels mounted on an aluminium support frame. Both ends of the cylinder are sealed to prevent gas leakage or infiltration during experiments. The speed of rotation of the cylinder can be varied between 2 and 10 RPM using a variable-speed motor. The slope of the dryer can be adjusted between 0° and 10°. A set of 9 removable lifters, 0.075 m wide and 0.6 m long, arranged in 3 series were equally distributed around the circumference, and bolted to the inside surface of the cylinder. Three 12-cm long bimetallic thermometers, installed at 0.8, 1.5 and 2.3 m from the feed end of the dryer, measure the gas temperature along the rotary cylinder with accuracy of 1°C. Three solids sampling ports are installed at 0.85, 1.55, 2.15 m from the feed end of the dryer to rapidly remove solids samples during experiments in order to measure its temperature and moisture content.

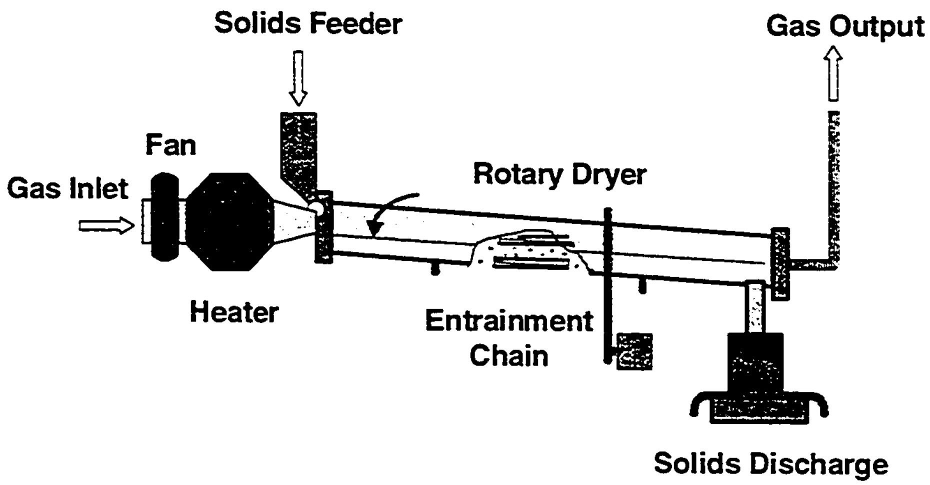

Figure 4.1 Schematic diagram of the pilot-scale rotary dryer.

An electric fan (WALMAR AXC315B) and an electric heater (GLENN ADH-30) were used to provide the desired gas flow rate and temperature, respectively. Under normal operating conditions, the gas flow rate will be in the vicinity of 10 kg/min at a temperature of approximately 120°C, which gives a superficial gas velocity of 2.6 m/s. The solids are continuously fed from the hopper to the rotary dryer using a revolving drum mounted at the base of the hopper. The required feed rate of solids up to 15 kg/min was set by increasing the rotation speed of the revolving drum and/or by adjusting the opening of a feeding gate

At the discharge end of the dryer, the solids were collected in a container placed directly on an electronic balance (OMEGA, LC2-12X12-100) and the mass recorded continuously with a microcomputer with accuracy 0.01 kg. The instantaneous discharge solids flow rate was determined by taking the time derivative of the accumulated mass of solids. The gas flow rate was measured with an anemometer (OMEGA HHF23) at the exit of the discharge duct. The drying gas was exhausted into the ventilation system.

Commercially available sand (Osgoode Sand & Gravel Company) that is widely used in the building industry was used as the solid material. The batch of sand was sieved to remove dust and particles having a diameter greater than 2 mm. The Sauter mean particle diameter of the dry sand was 444 μm and its bulk density was 1650 kg/m^3.

Experimental methods

The average residence time in an industrial rotary dryer can be estimated by imposing injecting an a pulse (delta-impulse) of a chemical tracer (LiCl, for example) directly at the entrance of the dryer and measuring its concentration at the discharge end [Renaud et al., 2000; Duchesne et al., 1996]. It is assumed that for the duration of the tracer experiment, the rotary dryer operates under steady state conditions. The impulse tracer test has the advantage to provide the residence time distribution in addition to the average residence time. However, such a type of experiment is difficult to perform in an industrial dryer mainly because of large amount of the tracer and complicated analysis of the tracer concentration. Apparently simpler experiments with solids tracers, such as marked particles, are hardly applicable in industrial conditions, especially when drying of pasty feeds and sticky materials. On the other hand, only the average residence time is often of prime interest to the engineers. If this value could readily be determined using a simple in situ method, it would be extremely helpful to adjust operating parameters of the dryer in order to maximize its performance. We postulate here that similarly to a step input tracer that is used in reactor engineering to determine the F curve [Levenspiel, 1972], a step-change in the input solids flow rate can be used to estimate the solids hold-up by integrating the area circumscribed between the step solids input ($t = 0$) and the discharge solids flow rate ($t = t_1$), as shown in Figure 4.2a. The same information can be obtained by making a negative step in the input solids flow rate, as shown at time t_2 in Figure 4.2b.

Once an estimate of solids hold-up, H, is available, the average residence time, τ, can be calculated from the hold-up with the following equation:

$$\tau = \frac{H}{F} = \frac{H_1}{F} = \frac{H_2}{F} \tag{1}$$

The solids hold-up was sometimes also physically measured in order to have an independent hold-up estimation to validate the simple solids input step method. When steady-state operation was achieved, the gas and solids flow rate, and the cylinder rotation were stopped suddenly and simultaneously. The solids were then removed section-by-section to determine both the solids hold-up distribution along the dryer, and the total hold-up. The solids flow rate F_i entering the dryer can be determined from the plateau in Figure 4.2. An additional measure of the average input solids flow rate was obtained by pre-weighting the total amount of sand fed to the dryer and recording the time when the last portion of solids enters the dryer.

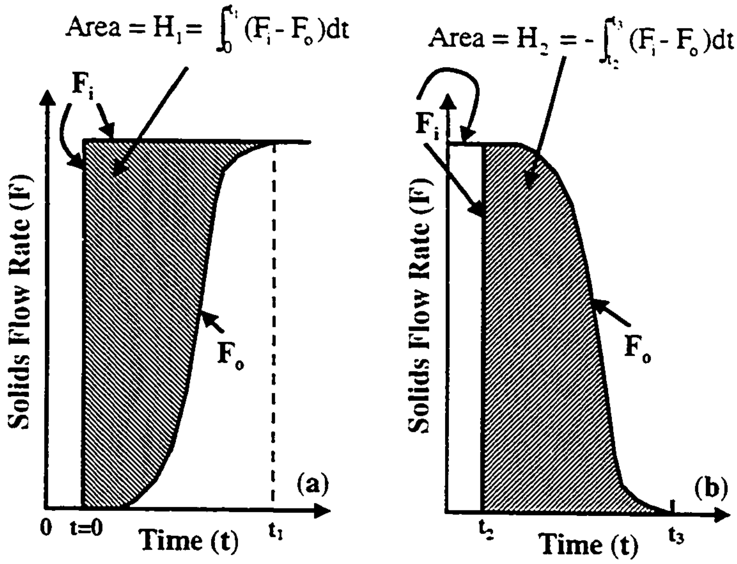

Figure 4.2 Dynamic step method used to estimate solids hold-up: (a) A positive step change in the input mass flow rate (F_i) from 0 to a constant flow rate and the output mass flow rate (F_o) is measured; (b) Similar to (a) for a negative step change in the input mass flow rate (F_i) from a constant value to zero.

EXPERIMENTAL RESULTS AND DISCUSSION

Dynamic determination of the solids hold-up and average residence time

Figure 4.3 shows a typical response of the discharge solids flow rate following a step change in the input solids flow rate from 0 to 6.7 kg/min at time zero, and a sudden interruption of the solids flow rate some time after a steady-state was achieved (about 37 min, in this experiment). At the beginning of the first step change, the dryer was completely empty. Integration of the surface between the input and the discharge flow rate, as depicted in Figure 4.2, provides two independent evaluations of the steady-state hold-up under given operating conditions. Several runs were performed under identical operating conditions with excellent reproducibility between runs as well as for the upward and downward steps in the solids flow rate during the same run (difference less than 3%). Occasionally, when steady-state operation was achieved, the dryer was stopped and the entire content of the dryer was removed and weighted to have an independent measurement of the hold-up and to validate the method proposed in this study. The total solids content determined by direct weighting and the hold-up estimated with the dynamic method were always very close to each other; the difference was no larger than 3%.

The inlet solids flow rate was evaluated using two distinct methods: (1) the steady discharge flow rate represented by the plateau in Figure 4.3, and (2) the time necessary for a pre-weighted amount of solids to flow from the hopper into the rotary dryer. With the estimations of the solids flow rate and the hold-up, the average residence time was calculated. Table 1 presents typical values of the hold-up, solids flow rates and the average residence time for a series of runs performed under different operating conditions.

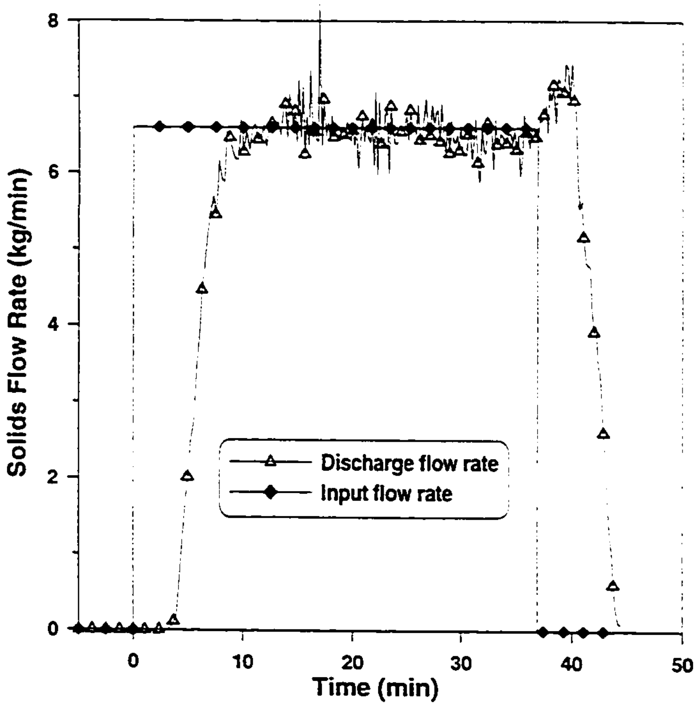

Figure 4.3 Typical solids discharge flow rate profile following step changes in the input solids flow rate.

The rotation speed of the rotary cylinder was controlled manually with a variable speed motor. The rotation speed was checked periodically during each run to maintain it as constant as possible. It was found that the speed of rotation had a tendency to slightly decrease when the hold-up increased and vice-versa when the hold-up decreased because the power necessary to rotate the cylinder increases with the mass of solids held in the dryer increases. This explains the transient increase in the solids flow rate observed in Figure 4.3 when the inlet flow rate was suddenly reduced to zero. To remedy this problem, an automatic rotation speed controller could to be installed. However, this artefact has no influence on the evaluation of the solids hold-up as it does not change the total amount of solids discharging from the dryer nor the integration area that is used to calculate the solids hold-up and the average residence time.

Table 1 Typical results for the estimation of the average residence time by the dynamic method

Run	Flow Rate (kg/min)		Hold-up (kg)			τ
	Mass/time	Plateau	Upward step	Downward step	Measurement	(min)
1	5.40	5.41	27.83	-	27.74	5.14
2	5.34	5.38	29.93	-	29.95	5.57
3	7.93	-	54.63	56.13	-	6.98
4	8.03	-	33.05	33.48	-	4.14
5	5.55	-	15.78	15.48	-	2.82

This simple dynamic method can easily be used in industry to estimate hold-up in rotary dryer during startup and shutdown periods provided that the discharge flow rate can be measured with sufficient accuracy, and nearly steady-state operation prevails after startup or before shutdown. This dynamic method can also be used during normal operation to determine the difference in solids hold-up when one of the process variables is changed. Following the initial step change with an empty dryer to estimate the solids hold-up, the change in solids hold-up, when one process variable is changed, can be determined by simply summing the integrated area between the input and the discharge solids flow rates. This procedure was used in a series of tests that were done to measure the influence of the input solids flow rate, the gas flow rate, the rotation speed, and the slope of the dryer on the solids hold-up. Figures 4.4 and 4.5 present the dynamic profiles of the solids discharge flow rate when a series of steps in the solids input flow rate were made with no gas flow and a gas flow rate set at 10.8 kg/min.

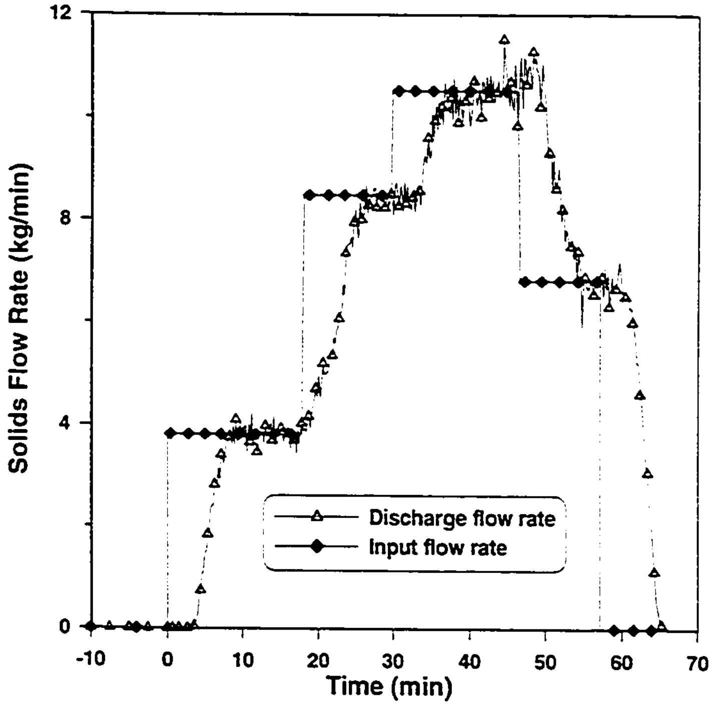

Figure 4.4 Discharge solids flow rate profiles following a series of step changes in the inlet solids flow rate (G = 0 kg/min).

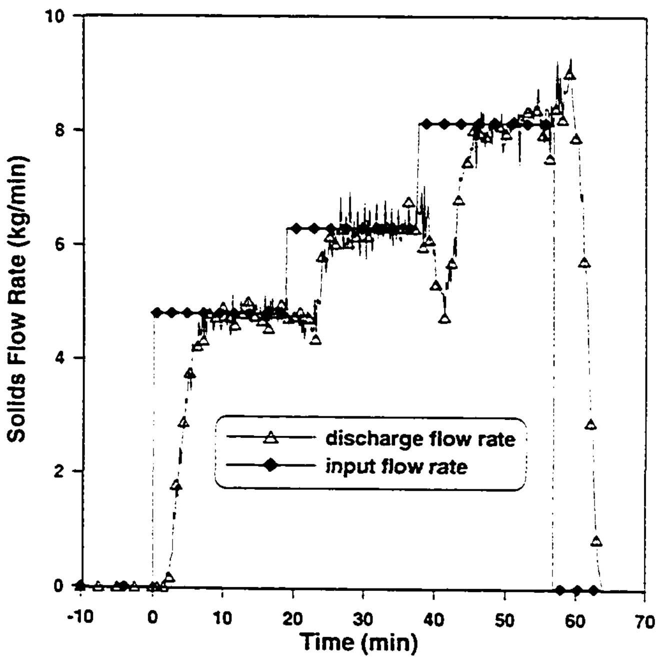

Figure 4.5 Discharge solids flow rate profiles following a series of step changes in the inlet solids flow rate (G = 10.8 kg/min).

The dynamic response of the discharge solids mass flow rate following a series of five step changes in solids flow rate are presented in Figure 4.4. By successive integration it is possible to determine the hold-up for each of the four different solids flow rates. The solids hold-ups given by the areas associated with these five steps were 21.0, 21.2, 10.7, -20.2, and -38.1 kg, from which the average residence time could be estimated using the cumulative area and the solids mass flow rate measured with the two methods described above. The summation of the five areas should theoretically lead to zero. For the experiment shown in Figure 4.4, the summation gave a residual of -5.4 kg, which is 4.9% of the summation of the absolute values of the five areas. This latter value corresponds to the maximum residual hold-up that was obtained in all the experiments. Table 2 presents the areas for the four experiments presented in Figures 4.4 through 4.7. The average residual of solids hold-up is approximately 3.0%. Even though the residual was relatively small, hold-up data reconciliation was performed by redistributing proportionally the residual hold-up to each area or the differential hold-up.

Table 2 Evaluation of solids hold-up

Figure	ΔH_1 (kg)	ΔH_2 (kg)	ΔH_3 (kg)	ΔH_4 (kg)	ΔH_5 (kg)	$\Sigma \Delta H_i$ (kg)	$\lvert \Sigma \Delta H_i \rvert / \Sigma \lvert \Delta H_i \rvert$ (%)
4.4	20.97	21.25	10.68	-20.15	-38.16	-5.41	4.86
4.5	19.73	8.724	14.43	-40.71	-	-2.17	2.60
4.6	48.66	-19.42	-11.23	6.88	-25.79	-0.90	0.80
4.7	28.38	-11.86	-8.52	4.19	-14.87	-2.68	3.95

Figure 4.5 presents similar results to those shown in Figure 4.4 with a gas mass flow rate of 10.8 kg/min. Very similar behaviour was observed except that the force exerted by the gas stream on the solids falling from the lifters leads to significantly lower hold-ups. This figure also shows the interaction of the gas stream on the solids. When the solids mass flow rate was raised to the highest value, the discharge flow rate decreased significantly at first and then increased to its steady-state value. It is hypothesized that the larger solids hold-up obtained at higher solids flow rates, results in larger mass of solids falling from the lifters, which reduces the impact of the gas mass flow rate on the solids transportation. This issue will be discussed later along with the relative influence of the solids and gas mass flow rate on the average residence time is discussed.

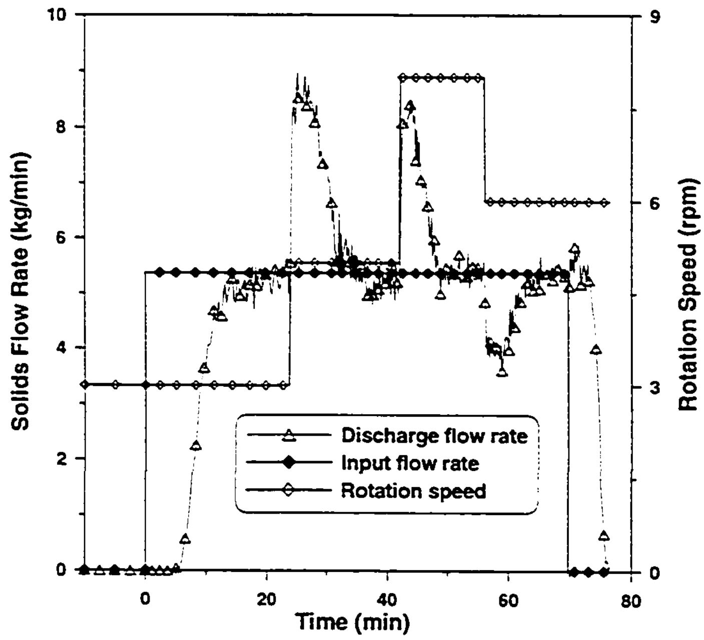

Figure 4.6 The dynamic response of the discharge flow rate following a series of step changes in the speed of rotation (G = 0 kg/min).

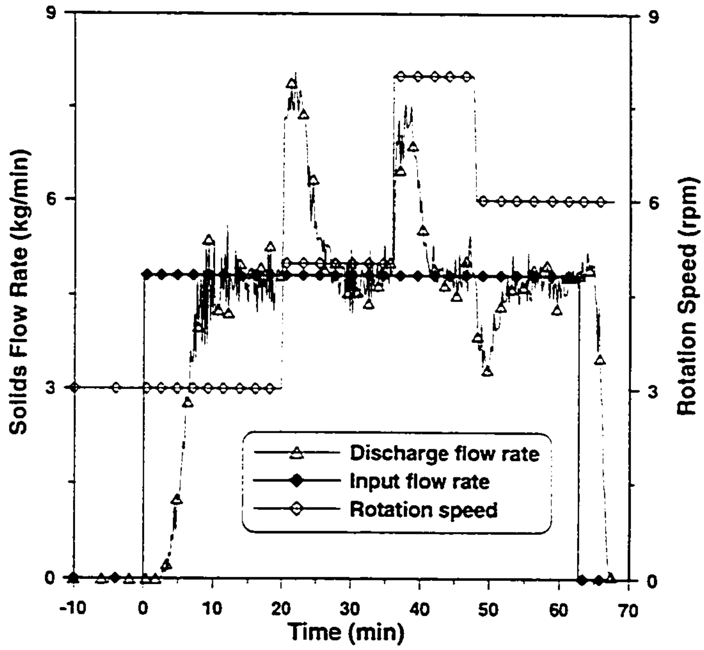

Figure 4.7 The dynamic response of the discharge flow rate following a series of step changes in the speed of rotation (G = 10.8 kg/min).

Figures 4.6 and 4.7 present the dynamic response of the solids discharge flow rate for step changes in the speed of rotation of the rotary dryer for the cases without and with a gas stream, respectively. It is interesting to observe that the speed of rotation has an immediate effect on the discharge solids flow rate. An increase in the speed of rotation will rapidly discharge a significant amount of solids before gradually returning to the new steady-state solids flow rate. A decrease in the speed of rotation leads to the inverse behavior. This distinctive and likely dynamic behavior that follows a step change in the speed of rotation, contrasts with the dynamic response for step changes of the other process variables, such as the solids and gas flow rates. The presence of a gas stream does not change significantly the dynamic response following a step change in the speed of rotation, except that the solids hold-up is smaller due to the drag force exerted on the particles that enhances their flow velocity through the cylinder.

Influence of the solids and gas flow rates on the average residence time

Figure 4.8 presents the variation of the average residence time in the rotary dryer as a function of the solids mass flow rate. For the type of solids used in this investigation and in the absence of a flowing gas, the average solids residence time is nearly constant over the entire range of solids mass flow rate. When the solids are in contact with a gas stream, the average residence time varies with the solids mass flow rate. For small solids mass flow rates, the drag force exerted on the solids by the flowing gas is more pronounced, and the particles are more easily entrained further down the cylinder in the direction of the flowing gas. The average residence time increases with the increase of the solids mass flow rate. The effect of the flowing gas decreases progressively as the solids flow rate, which is proportional to the solids hold-up, increases. At higher solids mass flow rates, the average residence time in the presence of the flowing gas approaches the average residence time when the flow rate of gas is zero. The gas is in contact with a larger quantity of solids and is not able to impart the same momentum to the cascading particles. This supports the hypothesis made when the results of Figure 4.5 were discussed.

The direct measurement of the local hold-up along the dryer showed that the solids hold-up was uniformly distributed. This behaviour is typical for free-flowing solids. Figure 4.9 presents the variation of the average residence time as a function of the gas flow rate for a constant solids mass flow rate (F = 5.13 kg/min). The average residence time decreases by half when the gas mass flow rate increases from zero to 12.3 kg/min. Small gas flow rates have a negligible influence on the solids average residence time. When the gas flow rate increases, a rapid decrease in the average residence time is noted.

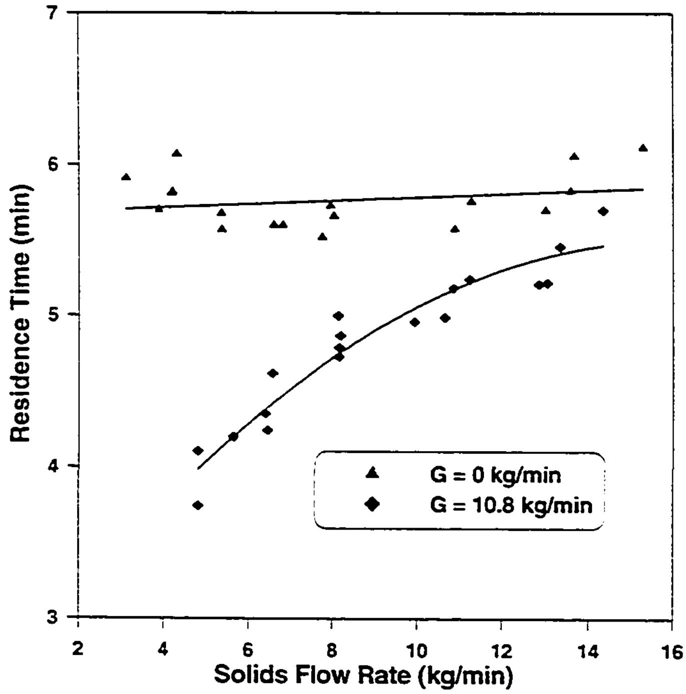

Figure 4.8 The variation of the average residence time with the solids flow rate.

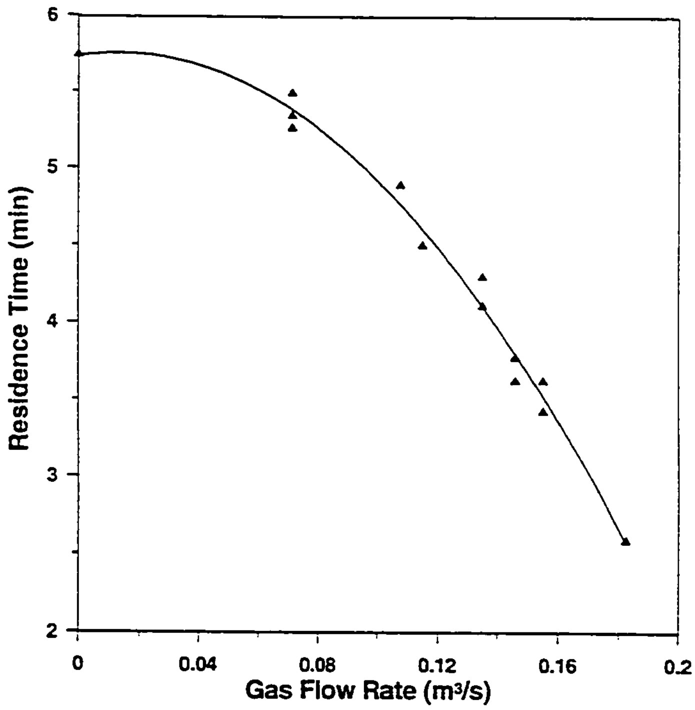

Figure 4.9 The variation of the average residence time with the gas flow rate (F = 5.13 kg/min).

Influence of rotation speed on residence time

As concluded from the results presented above, the rotation speed has an immediate action on the solids residence time. The solids transportation is a combined effect of the flights configuration that cascade the solids along the inclined cylinder, and the drag force exerted by the drying gas on solids particles. Figure 4.10 shows the variation of the average residence time with the speed of rotation without and with a gas mass flow rate of 10.8 kg/min. For both cases, a similar influence is observed. At small rotation speed, the residence time is high and a small change in the residence time produces a larger variation of the average residence time. At higher rotation speed, the average residence time approaches a limiting value where the speed of rotation ceases to have an effect.

Influence of dryer slope on residence time

Figure 4.11 shows the influence of the slope of the dryer on the average residence time. The average residence time varies almost linearly with the slope of the dryer. The influence of the gas flow rate on the average residence time is the same over the range of dryer slope studied. The main effect of the dryer slope is in the axial distance that solids are carried downstream during each cascade.

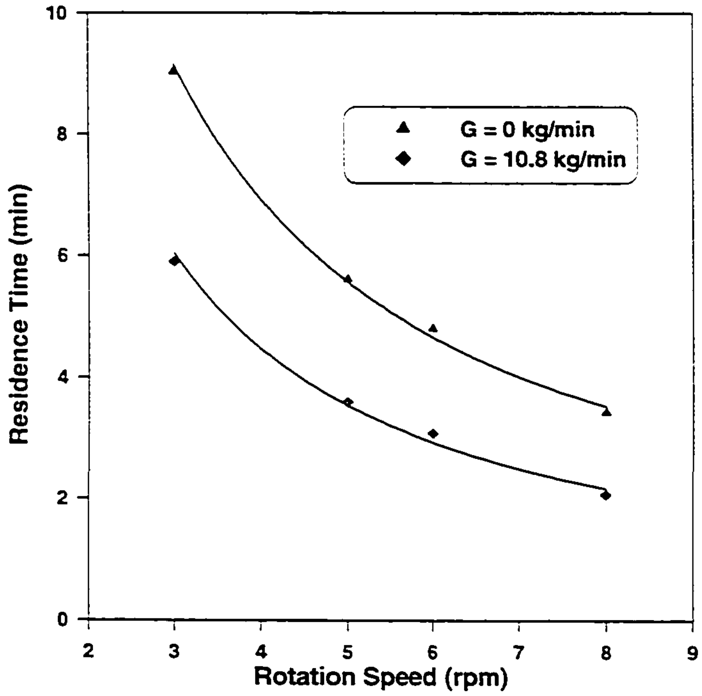

Figure 4.10 The influence of rotation speed on residence time.

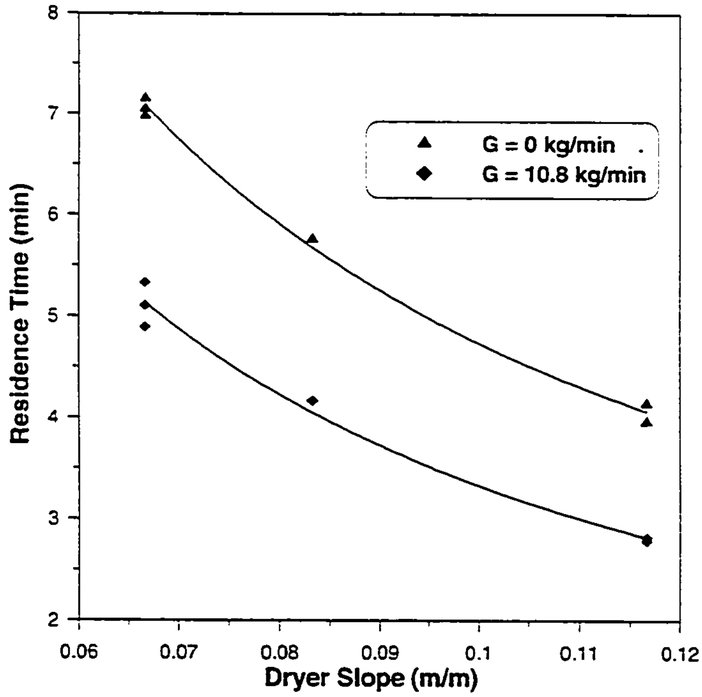

Figure 4.11 The influence of dryer slope on residence time.

Comparison of the mean residence time

Some correlations frequently used to estimate the average residence time are briefly described and evaluated using the series of data obtained in this investigation.

The first significant contribution to the estimation of the average residence time was proposed by Friedman and Marshall [1949]

$$\tau = \frac{10.58L}{sn^{0.9}D} \pm \left(\frac{608.3}{D_p^{0.5}}\right)\frac{LG}{F} \tag{2}$$

This equation accounts for the effect of the gas and solids mass flow rates as well as some characteristics and operating variables of the rotary dryer. The plus sign is for counter-current gas flow whereas the negative sign is for the co-current gas flow. This empirical equation was frequently used in the literature and is recommended for the design of rotary dryers. Foust et al. [1960] reported the above equation in a slightly modified form:

$$\tau = \frac{21L}{sn^{0.9}D} \pm \left(\frac{614.2}{D_p^{0.5}}\right)\frac{LG}{F} \tag{3}$$

Perry and Green [1984] cite equation 2 modified by increasing the constant in the first term of the equation by 30%, which corresponds to the average error that was observed by Friedman and Marshall [1949]

$$\tau = \frac{13.8L}{sn^{0.9}D} \pm \left(\frac{590.6}{D_p^{0.5}}\right)\frac{LG}{F} \qquad (4)$$

Figure 4.12 presents the comparison between the experimental residence times obtained in this study and the ones predicted using the above three correlations. This comparison clearly shows that the equation of Friedman and Marshall underestimates the average residence time whereas the correlation of Foust et al. overestimates the average residence time. The correlation given of Perry and Green is the most accurate for the system used in this investigation. Nevertheless, it appears that the slope of this correlation (Figure 4.12) is too high.

A revised correlation, using the identical structure of the previous correlations, was fitted using all the experimental results obtained in this investigation. Only four variables of this equation were varied: solids flow rate, gas flow rate, rotation speed, and slope of the rotary cylinder. The following correlation was obtained:

$$\tau = \frac{14.3L}{snD} - \left(\frac{354}{D_p^{0.5}}\right)\frac{LG}{F} \qquad (5)$$

The parity plot between the predicted and experimental average residence time is presented in Figure 4.13.

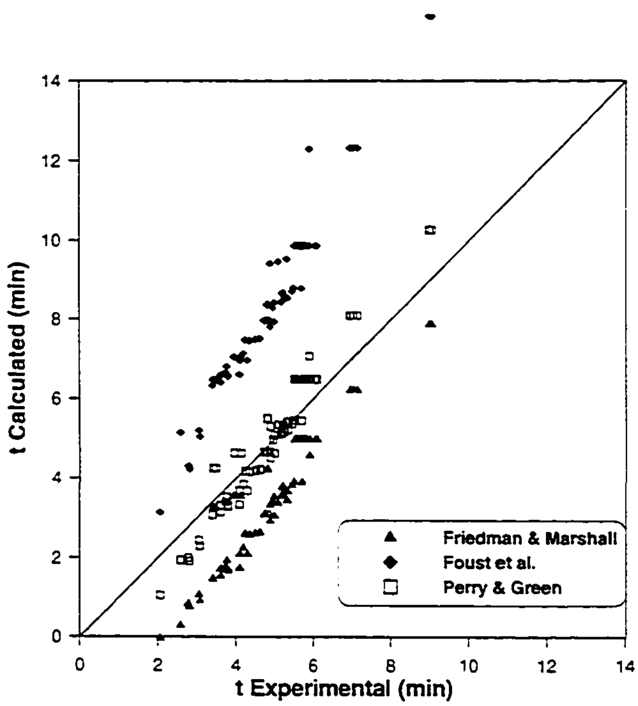

Figure 4.12 Comparison between experimental and predicted average residence time using equations 2, 3 and 4.

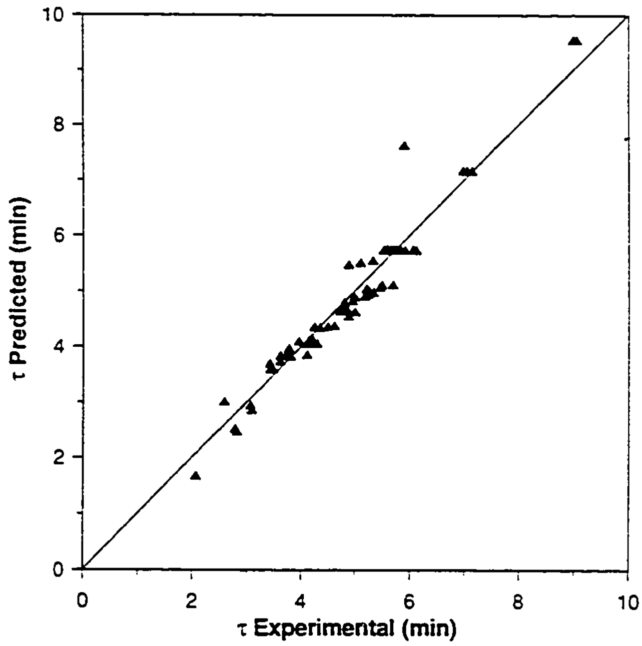

Figure 4.13 Comparison between experimental and predicted average residence time with equation 5.

A unit exponent for the rotation speed appears in equation (5) because the sum of squares of the errors was almost identical when the value of this exponent ranged from 0.8 to 1. The sum of squares of the differences between the predicted and experimental average residence time was reduced by a factor of four using equation (5) instead of equation (4). As the characteristics of solids have a significant impact on the ability of the particulate solids to move along the dryer, the constant of this correlation would probably be slightly different than those that would be obtained for another solid material in the same pilot-scale dryer or for the same solids processed in a different rotary dryer. In addition, some recent experimental results with wet solids in industrial and pilot-scale rotary dryers have clearly shown that the average residence time is greatly affected by the changing rheological behavior of the solids due to moisture content [Renaud et al., 2000; Renaud et al., 2001]. Therefore, the moisture content of solids and the characteristics of the solids need to be considered in order to develop a more general correlation. This is now being pursued in our laboratory.

SUMMARY AND CONCLUSION

This paper reports on the dynamic characteristics of solids transportation in a pilot-scale rotary dryer, and presents interesting results on the influence of the solids flow rate, gas flow rate, rotation speed and dryer slope on the solids average residence time. Results obtained clearly show that all these parameters have major impacts on the average residence time. Some correlations proposed in the literature to predict the average residence time were compared with the experimental values. In view of the excessive discrepancy, our own correlation for the mean residence time has been proposed.

The simple dynamic method based on the step-change in the solids feed rate has provided accurate and reproducible values of the mean residence time and solids hold-up. It is expected that the same accuracy and reproducibility would be achieved with other types of solids. For other dry or free-flowing solids, similar dynamic characteristics would undoubtedly be observed. Future experiments will be devoted to study the dynamic characteristics of wet solids. It is expected, however, that the simple dynamic method will be valid for these solids. The neat advantage of this simple dynamic response method is the possibility of determining in situ the influence of some keys variables such as flow rates and rotation speed on the average residence time.

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CHAPTER 5

SUMMARY AND CONCLUSIONS

Aside from the previous description of this work, some brief discussions related to the pilot-scale rotary dryer, solids transportation, and heat and mass transfer in rotary dryers are presented, followed by conclusions and recommendations for future works.

Pilot-scale rotary dryer

The pilot-scale rotary dryer designed for this work is a direct contact dryer with co-current flow between the hot flowing gas and the solids and is described in Chapter 4. The unit mainly consists of four parts: an electric fan, an electric heater, a solids feeding system and a rotary cylinder. Solids flow rate, gas flow rate, rotation speed and slope of cylinder could be adjusted in the range of 0 to 15 kg/min, 0 to 13 kg/min, 2 to 10 RPM, and 0° to 10°, respectively. The temperature and moisture content of solids along the rotary cylinder, the temperature of gas along the dryer and moisture content at the exit, the discharge amount of solids, and the temperature of outside surface of the cylinder can be measured during drying process and/or recorded continuously with a microcomputer.

Solids transportation

The knowledge of solids transportation is undoubtedly the most importation to modelling and design as it dictates the solids flow paths and residence time to contact the hot flowing gas inside rotary dryers. The mechanisms of solids transportation was examined in the Chapter 2, and the motion of solid particles flowing through a rotary dryer can have very irregular paths combining axial and radial motions. Solids transportation models were classified into seven models. Some experimental data with dry and moist solids in industrial and pilot-scale rotary dryers were used to study the influence of solids moisture content of on the solids average residence time, and to evaluate literature correlations. Results clearly indicate that the moisture content of solids has major impact on the solids residence time distribution (RTD) and on the average residence time. A second hump in the RTD was observed and it became predominant for high moisture solids content. Most of the literature correlation for the solids residence time is sufficiently accurate for dry and free-flowing solids whereas it much underestimates the residence time for solids with high moisture content. This is due to the changing rheological behaviour of the solids, typically having a varying degree of stickiness along the dryer.

A simple dynamic method was developed in this study to determine quickly and accurately the mean residence time and solids holdup in rotary dryer (Chapter 4). Numerous experiments were performed in a pilot-scale rotary dryer to study the dynamic characteristics of solids transportation, and the influence of some key operating variables on the mean residence time. Results definitely show that the solids flow rate, gas flow rate, rotation speed and dryer slope all have a significant impact on the average solids residence time. Some literature correlations to predict the mean residence time were compared and evaluated, and a slightly modified correlation was developed using all experimental results obtained in this investigation.

Heat and mass transfer

As the final purpose of the drying process is to get high quality and uniform products, the mechanisms of heat and mass transfer inside rotary dryers were also examined and solids transportation models were classified into four models (Chapter 3). Conductive, convective and radiative heat transfers all contribute, but in different proportions, to the drying process in rotary dryers. The influence of feed solids moisture content and gas temperature on the heat and mass transfer was investigated, and the volumetric heat and mass transfer coefficient were estimated and compared with some literature correlation. Results showed that inlet gas temperature had a major impact, with direct proportion, on drying rate, but the volumetric heat and mass transfer coefficients remained relatively constant. The mass transfer coefficients obtained from the heat transfer coefficients using the Chilton-Colburn analogy were very close to experimental values.

Recommendations

Based on all the work done during this thesis, the following recommendations for the future work are proposed:

- A study of solids transportation with wet solids should be carried out, incorporating the change in the residence time and solids holdup along the dryer. A new correlation should include new parameters to account for the characteristics of moist solids as a function of moisture content and of the nature of solids per se, in addition to the dryer characteristics and operating conditions. To do these experiments, a new feeding system should be designed in order to work for solids of all moisture content. In the current system, the feeding system was the bottleneck to study a wide range of solids moisture content.
- Investigations on the heat and mass transfer with dry and wet solid materials should be carried out further. In the current work, the total volumetric heat and mass transfer coefficients were estimated. Further studies should consider the change of local heat and mass transfer coefficients subjected to the change of moisture content of solids and solids holdup. More general correlations of heat and mass transfer coefficients, including the change of rheological behaviour of wet solids should be developed.
- Dynamic simulation of the total drying process in a rotary dryer, including solids transportation, and heat and mass transfer in steady

state and transient periods, should be done in order to develop new correlations.

APPENDIX A

A.1 Experimental results

Ori_F (kg/min)	Ori_G (m^3/s)	L (m)	D (m)	d_p (um)	ρ_p (kg/m^3)	F (kg/h m^2)	G (kg/h m^2)	n rpm	Slope (m/m)	τ (min)
3.10	0.0000	3.05	0.305	444	1650	2545.79	0.00	5	0.083	5.91
3.88	0.0000	3.05	0.305	444	1650	3186.34	0.00	5	0.083	5.70
4.20	0.0000	3.05	0.305	444	1650	3449.13	0.00	5	0.083	5.81
4.22	0.0000	3.05	0.305	444	1650	3465.56	0.00	5	0.083	5.82
4.32	0.0000	3.05	0.305	444	1650	3547.68	0.00	5	0.083	6.07
5.36	0.0000	3.05	0.305	444	1650	4402.57	0.00	5	0.083	5.68
5.37	0.0000	3.05	0.305	444	1650	4409.96	0.00	5	0.083	5.57
6.60	0.0000	3.05	0.305	444	1650	5420.07	0.00	5	0.083	5.60
6.81	0.0000	3.05	0.305	444	1650	5592.52	0.00	5	0.083	5.60
7.76	0.0000	3.05	0.305	444	1650	6372.68	0.00	5	0.083	5.52
7.96	0.0000	3.05	0.305	444	1650	6536.93	0.00	5	0.083	5.73
8.04	0.0000	3.05	0.305	444	1650	6600.16	0.00	5	0.083	5.66
10.88	0.0000	3.05	0.305	444	1650	8934.90	0.00	5	0.083	5.58
11.28	0.0000	3.05	0.305	444	1650	9260.92	0.00	5	0.083	5.76
13.01	0.0000	3.05	0.305	444	1650	10684.10	0.00	5	0.083	5.70
13.60	0.0000	3.05	0.305	444	1650	11168.62	0.00	5	0.083	5.83
13.68	0.0000	3.05	0.305	444	1650	11234.32	0.00	5	0.083	6.06
15.30	0.0000	3.05	0.305	444	1650	12564.70	0.00	5	0.083	6.12
4.81	0.1500	3.05	0.305	444	1650	3946.79	8702.16	5	0.083	4.11
4.82	0.1478	3.05	0.305	444	1650	3959.93	8574.53	5	0.083	3.75
5.64	0.1478	3.05	0.305	444	1650	4631.69	8574.53	5	0.083	4.20
6.40	0.1478	3.05	0.305	444	1650	5255.82	8574.53	5	0.083	4.35
6.45	0.1478	3.05	0.305	444	1650	5299.35	8574.53	5	0.083	4.24
6.57	0.1478	3.05	0.305	444	1650	5398.71	8574.53	5	0.083	4.62
8.14	0.1500	3.05	0.305	444	1650	6684.75	8702.16	5	0.083	5.00
8.15	0.1478	3.05	0.305	444	1650	6696.25	8574.53	5	0.083	4.73
8.16	0.1478	3.05	0.305	444	1650	6701.17	8574.53	5	0.083	4.79
8.19	0.1478	3.05	0.305	444	1650	6724.99	8574.53	5	0.083	4.87
9.94	0.1496	3.05	0.305	444	1650	8162.95	8678.96	5	0.083	4.96
10.65	0.1478	3.05	0.305	444	1650	8746.02	8574.53	5	0.083	4.99
10.85	0.1478	3.05	0.305	444	1650	8910.26	8574.53	5	0.083	5.18
11.24	0.1478	3.05	0.305	444	1650	9230.54	8574.53	5	0.083	5.24
12.86	0.1478	3.05	0.305	444	1650	10558.45	8574.53	5	0.083	5.21
13.06	0.1478	3.05	0.305	444	1650	10721.88	8574.53	5	0.083	5.22
13.36	0.1478	3.05	0.305	444	1650	10967.42	8574.53	5	0.083	5.46
14.35	0.1478	3.05	0.305	444	1650	11784.54	8574.53	5	0.083	5.70
5.76	0.0711	3.05	0.305	444	1650	4732.70	4124.83	5	0.083	5.26
6.94	0.0711	3.05	0.305	444	1650	5697.64	4124.83	5	0.083	5.49
5.66	0.0711	3.05	0.305	444	1650	4644.83	4124.83	5	0.083	5.34
5.48	0.1076	3.05	0.305	444	1650	4500.30	6242.35	5	0.083	4.89
5.05	0.1150	3.05	0.305	444	1650	4143.89	6671.66	5	0.083	4.50
4.86	0.1350	3.05	0.305	444	1650	3987.03	7831.95	5	0.083	4.11

Ori_F (kg/min)	Ori_G (m^3/s)	L (m)	D (m)	d_p (um)	ρ_p (kg/m^3)	F ($kg/h\ m^2$)	G ($kg/h\ m^2$)	n rpm	Slope (m/m)	τ (min)
4.86	0.1350	3.05	0.305	444	1650	3987.03	7831.95	5	0.083	4.30
4.42	0.1460	3.05	0.305	444	1650	3626.52	8470.11	5	0.083	3.62
4.97	0.1460	3.05	0.305	444	1650	4084.76	8470.11	5	0.083	3.77
4.58	0.1550	3.05	0.305	444	1650	3761.20	8992.24	5	0.083	3.42
4.93	0.1550	3.05	0.305	444	1650	4048.63	8992.24	5	0.083	3.62
4.04	0.1824	3.05	0.305	444	1650	3317.74	10581.83	5	0.083	2.59
6.11	0.0000	3.05	0.305	444	1650	5017.67	0.00	3	0.083	9.00
5.37	0.0000	3.05	0.305	444	1650	4409.96	0.00	3	0.083	9.06
5.36	0.0000	3.05	0.305	444	1650	4402.57	0.00	5	0.083	5.68
5.38	0.0000	3.05	0.305	444	1650	4414.89	0.00	5	0.083	5.57
6.11	0.0000	3.05	0.305	444	1650	5017.67	0.00	6	0.083	4.82
5.37	0.0000	3.05	0.305	444	1650	4409.96	0.00	6	0.083	4.80
5.36	0.0000	3.05	0.305	444	1650	4402.57	0.00	8	0.083	3.49
5.38	0.0000	3.05	0.305	444	1650	4414.89	0.00	8	0.083	3.43
4.81	0.1532	3.05	0.305	444	1650	3950.08	8887.81	3	0.083	5.90
4.93	0.1550	3.05	0.305	444	1650	4048.63	8992.24	5	0.083	3.80
4.58	0.1550	3.05	0.305	444	1650	3761.20	8992.24	5	0.083	3.42
4.82	0.1532	3.05	0.305	444	1650	3959.11	8887.81	5	0.083	3.80
4.81	0.1532	3.05	0.305	444	1650	3950.08	8887.81	6	0.083	3.09
4.86	0.1478	3.05	0.305	444	1650	3991.14	8574.53	6	0.083	3.07
4.82	0.1532	3.05	0.305	444	1650	3958.29	8887.81	8	0.083	2.07
8.76	0.0000	3.05	0.305	444	1650	7193.91	0.00	5	0.067	7.05
9.10	0.0000	3.05	0.305	444	1650	7473.12	0.00	5	0.067	7.15
7.93	0.0000	3.05	0.305	444	1650	6512.29	0.00	5	0.067	6.98
5.61	0.1496	3.05	0.305	444	1650	4607.06	8678.96	5	0.067	5.33
5.48	0.1496	3.05	0.305	444	1650	4497.01	8678.96	5	0.067	5.10
5.37	0.1496	3.05	0.305	444	1650	4413.25	8678.96	5	0.067	4.89
8.03	0.0000	3.05	0.305	444	1650	6596.88	0.00	5	0.117	4.14
8.03	0.0000	3.05	0.305	444	1650	6596.88	0.00	5	0.117	3.96
5.72	0.1496	3.05	0.305	444	1650	4695.42	8678.96	5	0.117	2.79
5.55	0.1496	3.05	0.305	444	1650	4556.14	8678.96	5	0.117	2.82