2

Basic Description of Rotary Kiln Operation

As seen in Chapter 1, unit operation equipment and other components are added together to form the rotary kiln system for material processing. Perhaps the most important is the rotary reactor, which forms the heart of any process and therefore warrants due attention. The rotary reactor is usually a long horizontal cylinder tilted on its axis. In most rotary kiln process applications, the objective is to drive the specific bed reactions, which, for either kinetic or thermodynamic reasons, often require bed temperatures that, for example, in cement kilns may approach as high as 2000 K. For direct fired kilns, the energy necessary to raise the bed temperature to the level required for the intended reactions, and in some instances, for example, the endothermic calcination of limestone, to drive the reactions themselves, originates with the combustion of hydrocarbon fuels in the freeboard near the heat source or burner. This energy is subsequently transferred by heat exchange between the gas phase (the freeboard) and the bed as was shown in Figure 1.1. Heat transfer between the freeboard and the bed is rather complex and occurs by all the paths established by the geometric view factors in radiation exchange (Figure 2.1). All these manifest themselves into a combined transport phenomenon with the various transport processes coming into play in one application. Often, because the analytical tools for handling freeboard transport phenomena have been the subject of considerable research. The ability

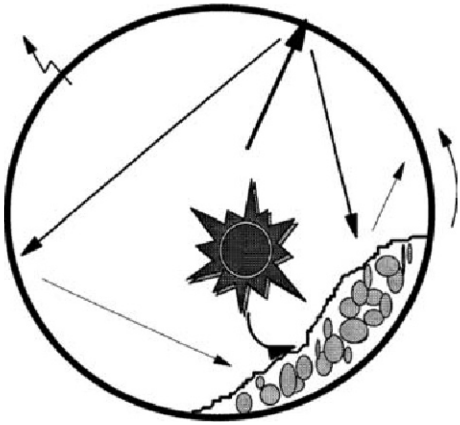


Figure 2.1 Radiation heat exchange in the cross section.

to simulate the freeboard conditions tends to exceed the ability to accurately determine conditions within the bed. For example, the zone method (Guruz and Bac, 1981) for determining radiative heat transfer, and commercial software for calculating fluid flow (and occasionally combustion processes as well) are well established. Thus the synergy between the freeboard phenomenon and the bed phenomenon, that is, the intended function of the process, gets distorted. One generalized model for estimating bed conditions is the well-mixed phenomenon, synonymous with the continuous stirred tank.

Numerous descriptions of the rotary kiln bed phenomenon have been proposed this way (Wes et al., 1976; Brimacombe and Watkinson, 1979; Tscheng and Watkinson, 1979; Gorog et al., 1981; and others). All of these assume that, at each axial position the bed is well mixed in the transverse plane, that is, the bed material is isothermal over any transverse section of the kiln. However many kiln operations suffer considerable difficulty in achieving a uniform product, one example being lime kilns, which experience chronic problems in preventing dead-burning of larger particles while fully calcining the finer particles. Evidence such as this, as well as operator experience, suggests that substantial transverse temperature gradients are generated within the bed. Thus, the well-mixed assumption, although expedient to the modeling of rotary kiln bed transport phenomena, is clearly deficient because one cannot ignore the motion of the bed in the transverse plane or the effect of this motion on the redistribution within the bed material of energy absorbed at the bed-freeboard interfacial surfaces.

In order to link the freeboard phenomenon to the bed phenomenon, the features of the bed should first be addressed.

2.1 Bed Phenomenon

During the thermal processing of granular materials in rotary kilns, heat transfer within the bed material occurs by the same mechanisms as in any packed bed, such as shaft kilns. Heat transfer paths at play can be particle-to-particle conduction and radiation, as well as interstitial gas-to-particle convection (Figure 2.2). However, the movement of the particles themselves superimposes an advective component for energy transport, which has the potential to dominate heat transfer. Hence the key feature of a rotary granular bed is the motion in the transverse plane, which sets the axial flow in motion and is dependent upon the rotation rate, degree of fill (volume of the kiln occupied by material) and the rheological properties of the particulate material.

Since the reactor is usually cylindrical and partially filled, generally it possesses two dispersion mechanisms, one in the axial direction that is characterized by an axial mixing coefficient, and the other in

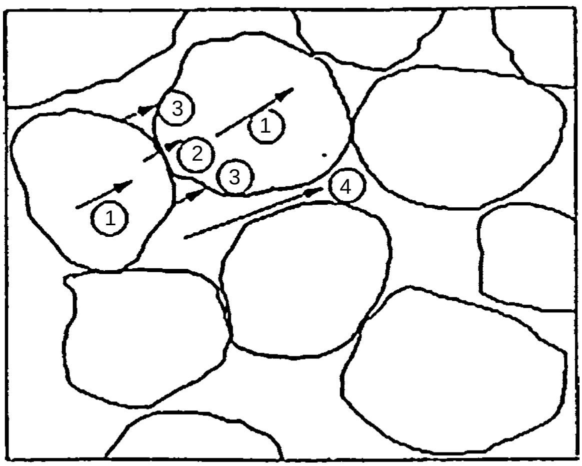


Figure 2.2 Bed heat transfer paths. 1. Internal conduction. 2. Particle-to-particle conduction. 3. Particle-to-particle radiation. 4. Interparticle convection.

the transverse direction, often associated with a radial mixing coefficient. The axial mixing component is attributed to the mechanism that results in an overall convection, and causes the bulk of the material to move from the inlet of the cylindrical drum to the outlet with an average velocity equal to the plug flow velocity. The radial mixing component, however, involves mechanisms at the smaller scale that cause local constraints on individual particles and result in velocity components both in the axial direction and the transverse direction. Both axial and transverse mixing coefficients tend to increase with an increase in kiln rotational speed. For low rates of rotation, one expects a spread in the residence time distribution due to the influence of the velocity profile (Wes et al., 1976). The effect of the drum size, particle rheology, and drum internal features are therefore major design considerations. The effect of the drum rotational speed on the transverse flow pattern is illustrated later. For now we present the features of the rotary reactor as a contactor by providing a quantitative description of the dispersion mechanisms, the resultant effects of which are critical to bed heat transfer during material processing.

2.2 Geometrical Features and Their Transport Effects

The key geometrical feature is the vessel size, given in terms of the cylinder diameter and kiln length related by the aspect ratio, that is, the length-to-diameter ratio (L/D) and also the slope. Other pertinent features include the internals, such as constriction dams and lifters, that impact the residence time. The empirical relationship developed by the US Geological Survey in the early 1950s relating the residence time and the kiln geometry has become a design mainstay even to this date (Perry, 1984). It is

$$\bar{\tau} = \frac{0.23L}{sN^{0.9}D} \pm 0.6\frac{BLG}{F} \tag{2.1}$$

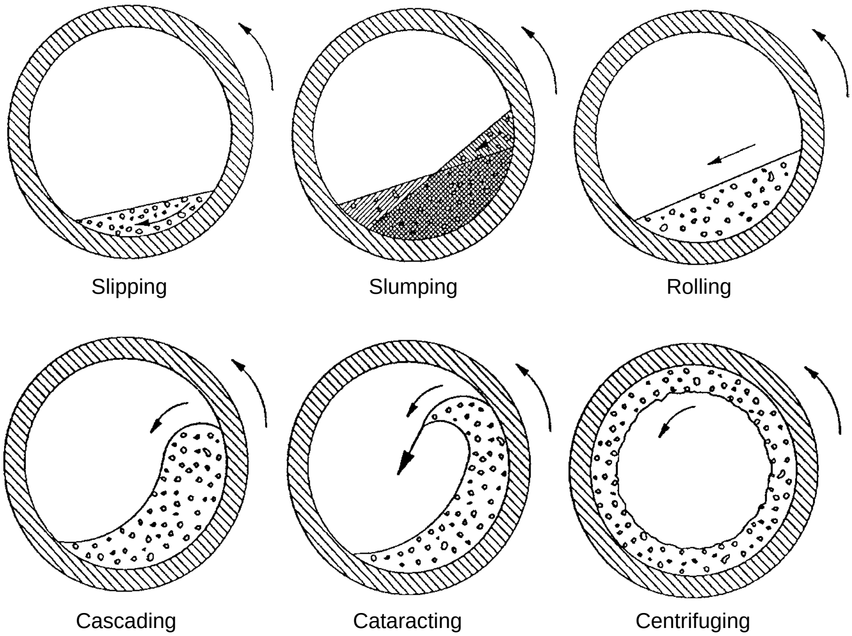
where s is the slope (ft/ft), N is the kiln rotational speed in revolutions per minute (rpm), L and D are the respective kiln length and diameter in ft, G is the freeboard gas velocity defined here in units of lb/hr/ft^2, and F is the feed or charge rate in lb-dry material per hour per square foot of cross sectional area. B is a constant depending on the material and is approximately defined as $B = sd_p^{-0.5}$ where d_p is particle size.

Because the vessel is partially filled and rotating on its horizontal axis, the freeboard or open space above the bed depends on the bed depth and, for that matter, on the kiln loading (% Fill). The shape of the free surface (the interface between the bed and freeboard) is dependent upon the operational requirements, that is, the feed rate, the drum rotational rate, and the material properties. As a result, the sizing of the rotary kiln depends on the application, typically, the feed rate (capacity) and related transport properties such as temperature, gas flow rates, and bed material velocities that ultimately will determine the residence time. For example, in dry processing applications, cylinder length-to-diameter ratios on the order of 5–12 are typical depending on whether the heat exchange is contact or non-contact. Such L/D ratios can result in residence times in the 20–120 minute range depending upon the kiln rotational speed, the type of internal flights, if any, and the slope in the longitudinal direction, typically in the range of 1–3°.

The movement of a charge in a rotating cylinder can be resolved into two components mentioned earlier, that is, movement in the axial direction, which determines residence time, and movement in the transverse plane, which influences most of the primary bed processes such as material mixing, heat transfer, and reaction rate (physical or chemical), as well as the axial progress of the charge. Although this linkage between particle motion in the transverse plane and particle velocity in the axial direction was established several decades ago, the literature generally deals with these two types of bed motion as independent phenomena until recent advances in the characterization and application of granular flow theories could be applied to powder processing in such devices (Boateng, 1998).

2.3 Transverse Bed Motion

Depending on the kiln's rotational rate, the bed motion in the transverse plane may be characterized as centrifuging, which occurs at critical and high speeds. This is an extreme condition in which all the bed material rotates with the drum wall. Cascading, which also occurs at relatively high rates of rotation, is a condition in which the height of the leading edge (shear wedge) of the powder rises above the bed surface and particles cascade or shower down on the free surface (Figure 2.3). Although operating the rotary kiln in either of these conditions is rare because of attrition and dusting issues, certain food



Mode	$Fr(=\omega^2R/g)$ at $\phi = 35°$
1. Slipping	$Fr < 1.0 \times 10^{-5}$
2. Slumping	$1.0 \times 10^{-5} < Fr < 0.3 \times 10^{-3}$
3. Rolling	$0.5 \times 10^{-3} < Fr < 0.2 \times 10^{-1}$
4. Cascading	$0.4 \times 10^{-1} < Fr < 0.8 \times 10^{-1}$
5. Cataracting	$0.9 \times 10^{1} < Fr < 1$
6. Centrifuging	$Fr > 1.0$

Figure 2.3 Bed motion in cross sectional plane. Froude numbers (*Fr*) are given for each of the different modes. (Henein, 1980.)

drying applications take advantage of the high particle-to-heat transfer fluid exposure associated with the cascading mode and the separation effect caused by the centrifugal force component. For example, starting at the other extreme, that is, at very low rates of rotation and moving progressively to higher rates, the bed will typically move from slipping, in which the bulk of the bed material, en masse, slips against the wall; to slumping, whereby a segment of the bulk material at the shear wedge becomes unstable, yields and empties down the incline; to rolling, which involves a steady discharge onto the bed surface. In the slumping mode, the dynamic angle of repose varies in a cyclical manner while in the rolling mode the angle of repose remains constant. It has been established (Rutgers, 1965) that the dynamic similarity of the rotary drum behavior, and hence the type of transverse bed motion that occurs during powder processing, is dependent upon the rotational Froude number, *Fr*, defined as

$$Fr = \omega^2 R/g \tag{2.2}$$

where the critical condition for centrifuging implies $Fr = 1$. The ranges of Froude numbers for the various modes are shown in Figure 2.3.

In the rolling mode (Figure 2.4), where rotary drum mixing is maximized, two distinct regions can be discerned, the shearing region, called the active layer, formed by particles near the free surface, and the passive or plug flow region at the bottom where the shear rate is zero. The particular mode chosen for an operation is dependent upon the intent of the application. A survey of various rotary drum type operations (Rutgers, 1965) has indicated that most operations are in the 0.04–0.2 range of N-critical, which is well below the centrifuging mode and probably the cascading mode as well.

The geometric features of a typical rolling bed are depicted in Figure 2.5. The bed is subtended at the continuous angle of repose ξ. The free surface is subtended at 2θ. Hence the bed cross section occupied by material can be defined by this angle. The chord length, L_c, the longest distance traveled by particles on the free surface (path of steepest decent), can also be defined in terms of this angle. The fraction of the cross sectional area occupied by material is the kiln loading. This is usually defined as the volume percent occupied by material in the

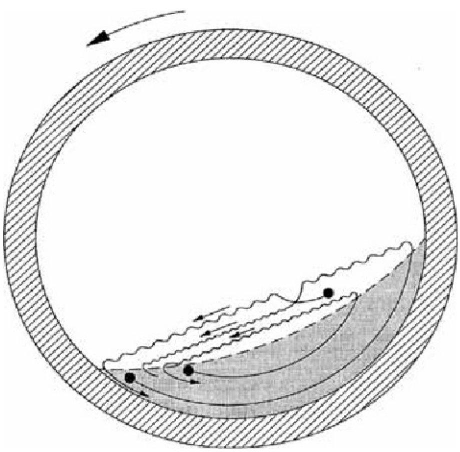


Figure 2.4 Rolling bed.

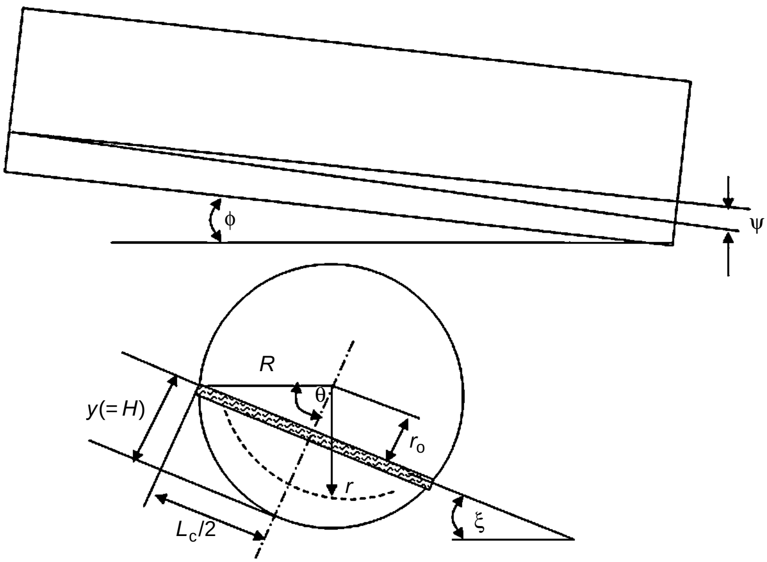


Figure 2.5 Rolling bed fill geometry.

vessel. Granted that the kiln length is constant, the degree of fill is the percent of the cross sectional area of the cylinder occupied by material (% Fill). The fraction filled defining the bed depth, and based on the geometry, relates the angles at any transverse section as follows:

$$f_c = \frac{1}{2\pi}\left(2\cos^{-1}\left(\frac{R}{R-H}\right) - \sin\left[2\cos^{-1}\left(\frac{R}{R-H}\right)\right]\right) \tag{2.3}$$

Seaman (1951) developed an approximation for the theoretical residence time of a shallow bed (lightly loaded kiln) and a theoretical relationship for the kiln volumetric flow rate for deep beds (heavily loaded kilns). Nonetheless, no clear definition has ever been given for the range of operation encompassed by the two cases of kiln loadings. Seaman's approximations led to the conclusion that kilns should be considered heavily loaded when the fractional cross sectional fill of solids exceeds approximately 5 percent.

As shown in Figure 2.4, two regions of the transverse plane can be discerned: (i) the active region near the top of the bed where surface renewal occurs, and (ii) the passive region beneath the active region. The active region is usually thinner than the passive region because particles there are not restricted and they move faster. Because the bed is constrained within the cylinder's geometrical domain, the laws of conservation of mass (i.e., particles going into the active layer must come from the plug flow region) requires that the depth of the active region be lesser than that of the passive region. This is due to higher particle velocity there, although there is more to this phenomenon than simple mass conservation. Most of the mixing in the drum cross section and also dissociation reactions occur in the active region. The deeper the active layer, the better the mixing. In order to increase the active layer depth, it is essential to increase kiln speed. Tracer studies have shown that particles move forward along the kiln only through the active region (Ferron and Singh, 1991). The kiln's longitudinal slope is usually small (1/2–3/4 in. per foot which is equivalent to 3–4°) and far less than the material angle of repose (typically 36–40°). Hence forward motion cannot be sustained by the gravitational component of the stresses alone indicating that the kiln's axial slope assists this forward motion but does not drive it. Therefore, for every kiln revolution, the bed material makes several excursions (possibly 3 or 4) in the cross section, thereby resulting in an axial advance. Increasing the kiln speed will result in an increase in the number of excursions and ultimately in increased mixing. Kiln speed increase, however, will decrease the residence time of the material, since the bed will move faster axially. It is therefore important for the kiln operator to know, based on the nature of the application, what the critical residence time should be in order to achieve the desired product quality. Process control by kiln speed is therefore critical if one can provide adequate mixing and maintain sufficient residence time to process the material.

Studies involving rotary kiln bed behavior have resulted in many ways of estimating how the bed will behave at any given operational condition. One such tool is a bed behavior diagram (Henein, 1980) which presents a typical behavior for a sand bed for a 41 cm (1.35 ft) diameter pilot kiln (Figure 2.6). Given the angle of repose, kiln geometry, and speed, users of such diagrams can predict what bed behavior to expect within the kiln cross section. However, since these delineation curves were generated from room temperature experiments, their industrial use has been limited. Despite this shortcoming, the bed behavior diagram can be helpful for the drying zone and, for the most part, the preheating zone of long kilns. In the calcination zone, however, any softening of the material that increases stickiness or agglomeration of the material will result in an increase in the angle of repose. Because of the importance of mixing on product uniformity, we will elaborate further on rolling bed behavior.

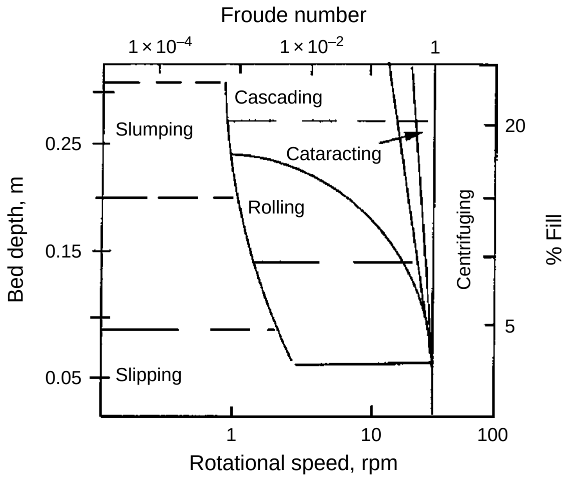


Figure 2.6 Bed behavior diagram. Mapping of bed behavior regimes for different operation conditions. (Henein, 1980.)

2.4 Experimental Observations of Transverse Flow Behavior

The key step in almost all particulate processing applications is solids mixing. Mixing is primarily used to reduce the non-uniformity in the composition of the bulk to achieve uniform blending or as a first step to improve the convective/advective and diffusion components associated with heat transfer to a particulate bed in thermal processing. Some of the observed flow and transport phenomena in rotary kilns that have provided insights into particulate flow behavior, which lead to accurately stating the mathematical problem or modeling the rotary kiln transport phenomena, are described. Dependent upon the bed depth and the operational conditions, the flow behavior constrained in the transverse plane can be purely stochastic, purely deterministic, or a hybrid of both; hence mixing can either be modeled by a random walk for very shallow beds (Fan and Too, 1981) or by a well-defined bulk velocity profile estimated by shear flows similar to boundary layer problems (Boateng, 1993). Early workers used tracer particles to observe and characterize mixing (Zablotny, 1965; Ferron and Singh, 1991; and others). Lately, such works have been extended to the use of nonintrusive techniques, such as nuclear magnetic resonance (NMR; Nakagawa et al., 1993) and positron emission particle tracking (PEPT; Parker et al., 1997).

For bulk processing, fiber-optic probes have been used in the past to establish bulk velocity profiles that have allowed estimates of the parameters necessary for bulk convection heat transport (Boateng, 1993). Boateng used experiments on the continuous flow of granular material in the transverse plane of a rotating drum to elucidate the rheological behavior of granular material in rotary kilns. Additionally, for the first time such work provided data for mathematical modeling of granular flow in the boundaries of a rotary kiln. Using granular materials that varied widely in their physical properties, specifically, polyethylene pellets, long grain rice, and limestone, the mean depth and surface velocities were measured using optical fiber probes (Boateng, 1993; Boateng and Barr, 1996). From these, granular flow behavior including the velocity fluctuations and the linear concentration of particles could be computed. The rotary drum used consisted of a steel cylinder of 1 m diameter and 1 m length. It had a glass end piece with a center opening providing access for flow measurements.

In light of the shape and thickness of the active layer, experiments confined to the observation of the general behavior of material flow in the cross section could be carried out. It was found that the transition of particles from the plug flow region of the transverse plane, where material moves in rigid motion with the cylinder, to the active layer, where particles are continuously shearing, not only depends on the material's angle of repose but also on the physical properties (such as the coefficient of restitution of the material). Because shear stresses generated in the induced flow involve collisional elasticity, particles with a relatively low angle of repose, for example, polyethylene pellets, experience an easy transition from the potential energy position in the plug flow region to kinetic energy in the active layer. Conversely, low coefficient of restitution materials, such as rice and limestone, have a relatively high friction angle and the energy dissipation is lower than that for polyethylene pellets. As a result, high potential energy is built up during the transition and is accompanied by material buildup prior to release into the active layer. Flow instabilities result, which manifest themselves into formation of multiple dynamic angles of repose with unsteady velocity distribution at the exposed bed surface.

The active layer depth and bed flow properties depend on the coefficient of restitution of the material. The flow properties of interest include granular temperature, which is a measure of kinetic energy in random motion of particles, and dilation. Granular temperature was found to be high at regions of low concentration with high mean velocity. These experiments also characterize the shape of the active layer to be parabolic irrespective of the material used. However, its thickness depends on the physical properties of the material and the operational parameters such as rotational rate.

After transferring from the plug flow region into the active layer, particles accelerate rapidly up to around mid-chord of the surface plane before decelerating with streaming, kinetic, and gravity effects playing various roles in momentum transfer. Increasing the degree of fill provides a longer chord length for material to travel and, for larger kilns, the velocity at the exposed bed surface can become fully developed by mid-chord. For deep beds, surface velocities can reach as high as 4.5–7.5 times the drum speed. Similar observations have been observed using nonintrusive flow measurement techniques (Parker et al., 1997), reproducing some of the results obtained by Boateng (1993).

2.5 Axial Motion

Axial dispersion is dependent upon transverse dispersion. For each excursion within the cross section, a particle on the free surface traveling on the chord length makes an axial move either backward or forward. This is due to the random nature of flow within the active layer. The forward advance is based on the cylinder slope in the axial direction as well as the apparent forward angle resulting from the transverse flow pattern. Some of the geometric relationships to the residence time for particles traveling from one end of the kiln to the other have been derived by various investigators since Seaman's work (1951). Although these derivations lack the rigor of a true granular flow, they have combined key empirical relationships to provide design formulas still in use today. Nicholson (1995) assembled some pertinent formulas that have been used to characterize axial bulk movement in kilns that are purely based on geometrical considerations. The major ones are from the early works by Seaman who calculated the material flow properties in the axial direction based purely on geometry of the equilibrium position (Figure 2.5).

For a small kiln slope ϕ, and the dynamic angle of repose ξ, the expression for the axial movement per cascade z_o, is given as

$$z_o = \frac{L_c\,(\phi + \psi \cos \xi)}{\sin \xi} \tag{2.4}$$

where ψ is the bed angle relative to the axial plane (Figure 2.5).

The number of cascades per kiln revolution mentioned earlier, N_c, might be estimated as:

$$N_c = \frac{\pi}{\sin^{-1}\left(\frac{L_c}{2r}\right)} \tag{2.5}$$

where r is the radius of the particle path in the bed. Multiplying the axial transport distance per cascade by the number of cascades per revolution and the kiln rotational speed yields the average axial transport velocity, u_{ax}, for a particle at position r anywhere in the radial plane.

$$u_{ax}(r) = nL_c\left(\frac{\phi + \psi\cos\xi}{\sin\theta}\right)\left(\frac{\pi}{\sin^{-1}\left(\frac{L_c}{2r}\right)}\right) \tag{2.6}$$

where n is the kiln rotational speed. For lightly loaded kilns (10–12 percent fill or less) one can make the following approximation

$$\sin^{-1}\left(\frac{L_c}{2r}\right) \to \frac{L_c}{2r} \tag{2.7}$$

The mean axial transport plug-flow velocity, u_{ax}, at a point along the kiln length can be calculated by the expression

$$u_{ax} = 2\pi rn\left(\frac{\phi + \psi\cos\xi}{\sin\theta}\right) \tag{2.8}$$

The average residence time can be estimated from Equation (2.8) knowing the kiln length, L,

$$\bar{\tau} = \frac{L\sin\xi}{2\pi rn(\phi + \psi\cos\xi)} \tag{2.9}$$

For beds with a low degree of fill and no end constriction dams, one can practically assume a uniform bed depth throughout the entire kiln length ($y = H$) and, $\psi \to 0$. The residence time expression can be reduced to

$$\bar{\tau} = \frac{L\sin\xi}{2\pi rn\phi} \qquad \bar{\tau} = \frac{L\theta\sin\xi}{L_c\omega\tan\xi} \tag{2.10}$$

It can be easily recognized that Equation (2.10) resembles the empirical formula for the determination of the average residence time, Equation (2.1), established by the US Geological Survey.

One can also determine the maximum capacity of the kiln by the same geometrical considerations if the flow area can be estimated. From the geometry (Figure 2.5) the chord length can be expressed as

$$L_c = 2\left(r^2 - r_o^2\right)^{1/2} \tag{2.11}$$

$$u\mathrm{d}A = 4\pi n\left(\frac{\phi + \psi\cos\xi}{\sin\theta}\right)\left(r^2 - r_o^2\right)^{1/2} r\,\mathrm{d}r \tag{2.12}$$

Integrating Equation (2.12) from $r = r_o$ to $r = R$ gives an expression for the average volumetric axial transport rate as

$$q = \frac{4\pi n}{3}\left(\frac{\phi + \psi\cos\xi}{\sin\theta}\right)\left(r^2 - r_o^2\right)^{3/2} \tag{2.13}$$

For beds with dams, and therefore a nonuniform bed depth along the kiln length, the rate of change of bed depth can be estimated in terms of the subtended angles as

$$\frac{\mathrm{d}r_o}{\mathrm{d}z} = \frac{3q\sin\xi}{4\pi n\cos\xi\,(R^2 - r_o^2)^{3/2}} - \frac{\phi}{\cos\xi} \tag{2.14}$$

Nicholson (1996) found good agreement between such theoretical calculations and experimental data for cylinders with and without constrictions at the discharge end.

Several early and recent investigators have derived variations of the throughput, axial velocity, and residence time expressions (Hogg et al., 1973; Perron and Bui, 1990) but Seaman's expressions given here have found most practical applications and are recommended for use in industrial kilns so long as there are no true granular flow models to predict the bed behavior as are available for fluids of isotropic materials.

2.6 Dimensionless Residence Time

Using the same geometric considerations by Seaman, dimensionless residence time and flow rate may be derived (McTait et al., 1995) as a function of *f(y/R)*. Recognizing that this derivation might be convenient for estimating these parameters for all kilns. Using $L_c = y(2R - y)$ and $\sin\theta = L_c/R$, the dimensionless residence time can be expressed as

$$\bar{\tau}_d = \frac{\bar{\tau} nD\tan\phi}{L\sin\xi} = \frac{\sin^{-1}\left[(2 - y/R)\,y/R\right]^{1/2}}{\pi\left[(2 - y/R)\,y/R\right]^{1/2}} = f\,(y/R) \tag{2.15}$$

Similarly, using $A_s = R^2\theta - L_c(R-y)$ a dimensionless volumetric flow can also be expressed in terms of the bed depth, recognizing that $q\bar{\tau} = LA_s$, which is

$$q_d = \frac{q\sin\xi}{nD^3\tan\phi} = \frac{\pi}{4}\left[(2-y/R)\right]^{1/2} - \frac{\pi}{4}\frac{(2-y/R)(1-y/R)y/R}{\sin^{-1}\left[(2-y/R)y/R\right]^{1/2}} \tag{2.16}$$

and for the condition of uniform bed with no surface inclination in the axial direction, that is, $dy/dz \to 0$,

$$q_d = \frac{\pi}{6}\left[(2-y/R)y/R\right]^{1/2} \tag{2.17}$$

Although averaged or plug-flow estimations can be deduced from the geometry of the kiln using the equilibrium angles, the actual residence time can completely differ from these estimates. This is due to the random nature of solids motion alluded to earlier. For each cascade or excursion, a particle can move forward in line with the forward geometric projection as discussed, or backward due to several factors including changing material dynamic angle of repose. Evidence of this lies in the situation where large agglomerated particles (notoriously known in the industry as *logs*) which form in high-temperature kilns have been observed to travel back and forth without being discharged. Also in the cross sectional plane, particles in the plug flow region can emerge in the active layer in a sequence following the birth-death phenomenon, that is, entirely by random walk (Ferron and Singh, 1992). Hence residence time is a result of axial dispersion, which in turn depends on transverse dispersion and is truly a distribution function. It is not surprising, therefore, that residence time has been the subject of many tracer experiments. Parker et al. (1997) using the PEPT for particle imaging have shown that the distribution of axial displacement follows a Gaussian distribution. They have observed in their experiments a one-to-one relationship between the radial entry of a tracer particle and its exit within a bed of industrially processed materials.

This notwithstanding, the variance σ' over the average path traveled by particles on the free surface (some fraction over the chord length, L_c) might be included in the geometric derivations from an operational standpoint (McTait et al., 1995).

These relationships between residence time and bed depth can be plotted and used in tandem for kiln design. Given the diameter, kiln slope, and desired L/D, the residence time and the feed rate that will

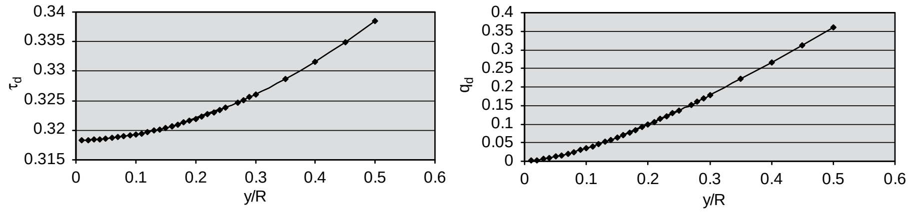


Figure 2.7 Dimensionless residence time and flow rate for 3.66 m diameter kiln with $L/D = 10$, slope = 3°, and material angle of repose = 40°.

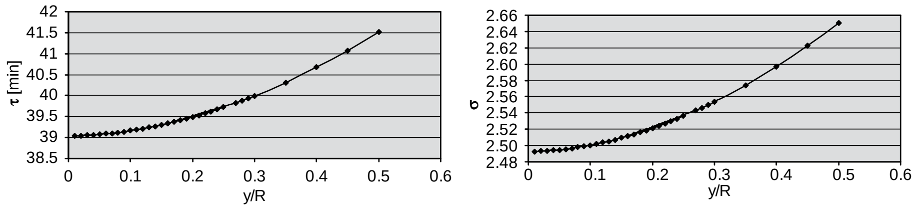


Figure 2.8 Calculated mean residence time and variance for 3.66 m diameter kiln with $L/D = 10$, slope = 3°, and material angle of repose = 40°.

result in a desired bed depth at a specified kiln speed can be calculated fairly well using these relationships. Combined with process data, these can be manipulated to achieve optimum kiln size and operating conditions. A plot of the dimensionless residence time and flow rate for a 3.66 m (12 ft) kiln, $L/D = 10$, a kiln slope of 3°, and 40° angle of repose operating at 1 rpm are shown in Figure 2.7. From these the actual residence time and related variance and also the maximum volumetric flow rate for a typical industrial material with 1600 kg/m^3 (100 lb/ft^3) bulk density can be estimated for design purposes (Figure 2.8).

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