Residence Time Distribution and Material Flow Studies in a Rotary Kiln

P.S.T. SAI, G.D. SURENDER, A.D. DAMODARAN, V. SURESH, Z.G. PHILIP, and K. SANKARAN

Experiments were conducted in a rotary kiln containing ilmenite particles to study the residence time distribution (RTD) of low-density particles, holdup, and bed depth profile. The variables include feed rate of solids, slope and rotational speed of the kiln, type and size of the tracer, and dam height. Correlations are presented for mean residence time, dispersion number, holdup, and steady-state throughput of solids in terms of the process variables. A simple method is proposed to estimate the dam height that gives rise to a flat profile of solids bed along the length of the kiln.

I. INTRODUCTION

HETEROGENEOUS noncatalytic gas-solid reactions play an important role in many extractive metallurgical and chemical processes. Rotary kilns continue to find extensive applications for such gas-solid reactions, despite challenges from newer and more specialized reactors. The typical applications include drying, heating, calcining, reducing, roasting, or sintering of a variety of materials. Successful modeling of such systems requires the knowledge of the importance of various parameters that influence the flow of materials and residence time distribution (RTD). Also, to simulate the kinetic behavior of a continuous reaction and to design a rotary kiln, it is required that the mixing behavior of material in the kiln be properly described.

Industrial processes which employ rotary kilns involve either a single type of solid, as in calcination, or two solids, such as in the direct reduction of iron ore or ilmenite. While a number of investigations have aimed at elucidating the solid mixing and material flow behavior for a single type of solid in a rotary kiln,[1,2,3] there has been hardly any work reported on the binary solid systems, which can have widely differing densities and particle sizes. For example, in the rotary kiln-based direct reduction process, ilmenite and coal are employed, having a density ratio of almost a factor of 3 and widely different particle sizes.

While a number of studies on rotary drums[4–7] (L/D ratios less than 10) have been reported, investigations on rotary kilns have been relatively scarce. Investigations on the solids behavior in rotary kilns essentially have three distinct objectives, namely, material flow and holdup characterization, radial segregation of solids with respect to the size differences, and regimes of solid motion in the transverse direction. Chatterjee *et al.*,[1,2,3] based on their detailed investigation, have reported the effect of kiln operating parameters and geometry on the averaged residence time of solids, holdup, and kiln throughput. They also investigated the effect of ring formation within the kiln. The authors proposed a correlation for residence time of the charge. Among the operating parameters, the effect of exit dam height was not reported. Henein *et al.*[8] made a detailed investigation on the effect of rotational speed and diameter of the kiln, size and type of the particles, and bed depth on bed motion and proposed bed behavior diagrams to delineate the various regimes, namely, slumping, transition, and rolling. In a subsequent paper,[9] the authors developed a mathematical model to predict the conditions giving rise to the different forms of transverse bed motion in rotary kilns. While analyzing the ilmenite reduction reaction, Sai *et al.*[10] have reported the significant influence of solid-solid mixing on the overall reduction.

Recent publications[11,12] on the mechanism of solids flow in rotary kilns have focused on the radial segregation of particles of different sizes. These studies have elucidated the basic mechanism for radial segregation, namely, percolation, and flow of solids with the use of three different sizes of solids. They were successful in observing the segregation kinetics by high-speed photography, on the basis of which they proposed a zero-order rate of segregation. Correlations were also proposed to relate the specific segregation rate constant with the Froude number at different ratios of particle sizes.

When solids undergo chemical conversion, a critical aspect would be the residence time of the solids in the rotary kiln. The work on radial segregation clearly indicates that the residence time of the solids would depend on the size and density differences of the solids. The measurement of RTD by the tracer technique is a powerful tool to characterize the influence of characteristics of solids on their residence time in the reactor.

This paper reports a detailed investigation on the RTD of low-density solid tracer particles (coal and sand) in a bed of high-density solids (ilmenite) carried out in a rotary kiln, and tracers are supposed to mimic the motion of lighter particles in the presence of heavier particles. The RTD curves were obtained and evaluated in terms of the dispersed plug flow and tanks-in-series models. The corresponding dispersion number and the equivalent number of tanks in series are estimated for different operating variables. Correlations are proposed for mean residence time and dispersion number in terms of the operating variables. The solid tracers have been selected

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such that their size range is larger than the bed material (ilmenite) but having lower densities. In this manner, the movement of the solid tracer through the bed would be only by kiln action. Percolation effects will be absent.

Bed heights were measured at different lengths from the exit end of the kiln, and the relation between throughput of solids and operating variables was observed and correlated under steady-state conditions. The concept of "kiln carrying capacity" is explained, and a method to estimate the dam height that is required to maintain a flat profile of filling of solids is proposed.

II. EXPERIMENTAL APPARATUS AND METHOD

Experiments were conducted in a rotary kiln covering a wide range of variables, as detailed in Table I. Table II lists the physical properties of the working material and the tracers used. Figure 1 shows the schematic diagram of the experimental setup. The kiln used had an internal diameter of 147 mm and a length of 5900 mm, giving an L/D ratio of 40. The kiln was fabricated from 6-mm-thick 316 stainless steel plate. A screw feeder was used to feed the material to the kiln at a constant rate. To drive the kiln at the desired rotational speed, a direct current (DC) motor was provided, while the slope of the kiln was varied with the help of a jack mechanism provided at the feed end of the kiln. Circular dams were used at the exit end with heights of 15 and 28 mm, thereby covering 20 and 38 pct inside diameter of the kiln, respectively. The dam at the feed end is a fixed one of 30-mm height.

All experiments were conducted at room-temperature conditions. Ilmenite ore of $-0.3 + 0.1$ mm size was used as the working material. Under steady-state conditions, the RTD was determined by introducing the tracer as a pulse through the screw feeder. This was effected by replacing the normal feed for 1 minute with the tracer. However, correction was introduced in time for the tracer particles to reach the beginning of the reactor tube from the feed hopper. Samples were collected at regular intervals at the discharge end of the kiln for a known amount of time. The tracer was then separated from the sample manually, and the weights of the tracer and ilmenite were noted. The quantities of tracers used were: sand = 570 g and coal = 245 g.

The bed depth at different axial lengths along the kiln was measured using a rod and pointer arrangement. The pointer was oiled by means of lubricating oil, and the rod was inserted into the kiln. The pointer was then made to touch the bottom of the kiln tube through the bed. The rod was then taken out, and the length of ilmenite that was stuck on the pointer was measured to give the bed depth at that point (Figure 6). This procedure was repeated for different axial distances of 400, 650, 1000, and 1750 mm from the discharge end.

The holdup was determined by stopping the kiln after the steady state was attained and subtracting the total quantity of solids that was collected from the exit end from that fed to the reactor. The throughput of solids

Table II. Physical Properties of Tracers and Working Material

Solid	ρ (kg/m^3)	R (deg)
Ilmenite	4200	27.4
Sand	2500	21.8
Coal	1600	24.6

Table I. Details of Experimental Conditions of the Present Study

Run Number	Tracer Type	Tracer $d_p \times 10^3$ (m)	H (m)	θ (deg)	N (rpm)	F (kg/h)
1	sand	1.1	0.015	1.37	3	36.0
2	sand	1.1	0.015	1.37	3	28.0
3	sand	1.1	0.015	1.37	3	13.3
4	sand	1.1	0.015	1.37	1	13.3
5	sand	1.1	0.015	1.37	1	6.8
6	sand	1.1	0.015	1.37	2	13.3
7	sand	1.1	0.015	0.78	3	13.3
8	sand	1.1	0.015	1.10	3	13.3
9	sand	1.1	0.028	1.37	3	28.0
10	sand	1.1	0.028	1.37	2	13.3
11	sand	1.1	0.028	1.37	2	6.8
12	coal	1.1	0.015	1.37	3	36.0
13	coal	1.1	0.015	1.37	3	28.0
14	coal	1.1	0.015	1.37	3	13.3
15	coal	1.1	0.015	1.37	1	13.3
16	coal	1.1	0.015	1.37	1	6.8
17	coal	1.1	0.015	1.37	2	13.3
18	coal	1.1	0.015	0.78	3	13.3
19	coal	1.1	0.028	1.37	3	28.0
20	coal	1.1	0.028	1.37	2	13.3
21	coal	1.7	0.015	1.10	3	13.3
22	sand	1.7	0.028	1.37	2	6.8
23	sand	2.4	0.028	1.37	2	6.8

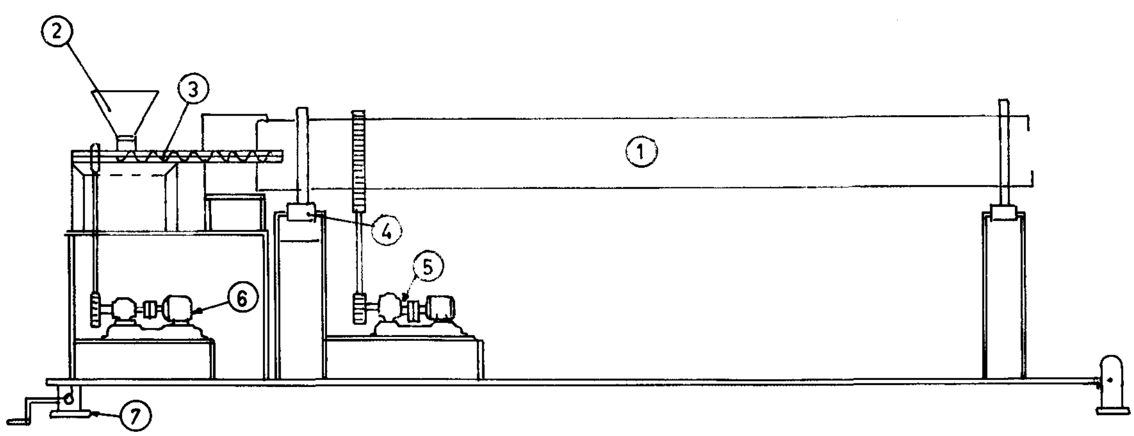

Fig. 1—Schematic of the experimental setup: (1) kiln tube; (2) hopper; (3) screw feeder; (4) roller; (5) kiln tube drive; (6) screw feeder drive; and (7) lifting jack.

and the bed depths under dynamic conditions were also noted.

III. RESULTS AND DISCUSSION

A. *Residence Time Distribution*

Several preliminary experiments were conducted to test the experimental apparatus and the reproducibility of results. Under similar conditions, the variation in mean residence time was within ±5 pct, thus establishing the reproducibility of the results. A set of 23 experiments was performed for the purpose of determining the mean residence time of particles having different physical characteristics and for different operating conditions. The experimental conditions used in each run are summarized in Table I. Runs 21 through 23 are not considered for obtaining the correlations.

A typical tracer output signal is shown in Figure 2, plotted as normalized concentration *vs* time; the normalization was done by referring the tracer concentrations to the total amount of tracer introduced.

It is sometimes recommended[13] that the best manner of comparing RTDs is by using their moments, instead of trying to compare the entire distribution. For this purpose, three moments are normally used. The first of these is the mean residence time. The second moment commonly used is called the "variance" or "square" of the standard deviation. The magnitude of this moment is an indication of the spread of the distribution; the greater the value of this moment, the greater the distribution's spread. The third moment is also taken about the mean and is called "skewness." The magnitude of this moment measures the extent that the distribution is skewed in one direction or another in reference to the mean.

The *C*-curve, the effluent concentration-time curve, is converted to the *E*-curve, the RTD curve, by

$$E(t) = C(t) \Big/ \int_0^\infty C(t)\,dt \qquad [1]$$

The mean residence time is then calculated by

$$T_e = \int_0^\infty tE(t)\,dt \qquad [2]$$

The area under the curve of a plot of $tE(t)$ as a function of t yielded T_e, calculated by Simpson's three-eighths rule.[13]

The variance of residence times about the mean residence time is then calculated by

$$\sigma^2 = \int_0^\infty (t - T_e)^2 E(t)\,dt \qquad [3]$$

The area under the curve of the plot of $(t - T_e)^2\,E(t)$ as a function of t, calculated in a similar manner to T_e, yielded σ^2.

The variance is converted to dimensionless variance by

$$\sigma_\theta^2 = \sigma^2 / T_e^2 \qquad [4]$$

The analysis of the present system was carried out by considering the monodimensional models,[14] *i.e.*, the dispersion model and tanks-in-series model. The vessel

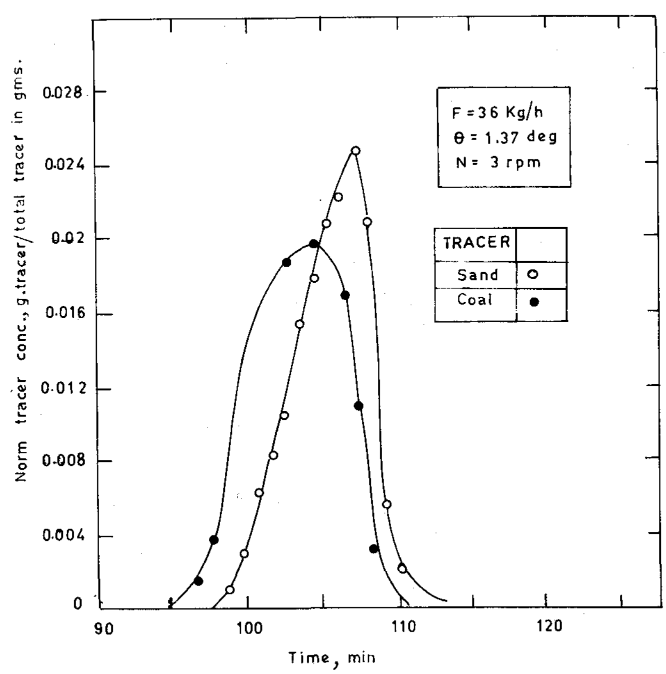

Fig. 2—Typical tracer output signal.

dispersion number can be calculated from the experimental results as a function of the variance by the following relationship:[14]

$$\sigma_\theta^2 = 2(D/UL) - 2(D/UL)^2[1 - \exp(-UL/D)] \quad [5]$$

In the present study, since (D/UL) values are very small, the above equation can be approximated as

$$\sigma_\theta^2 = 2(D/UL) \quad [6]$$

In the case of the tanks-in-series model, the number of equal sized back-mix flow tanks (n) is estimated by

$$\sigma_\theta^2 = 1/n \quad [7]$$

The mean residence time, the dispersion number, and the number of tanks thus estimated are listed in Table III.

As can be seen from Table III, the mean residence time increases with increase in dam height but decreases with increase in either rotational speed or inclination of the kiln. This observation is in agreement with that reported by Chatterjee *et al.*[2] On the other hand, it is independent of feed rate of solids and tracer.

The mean residence time thus computed is correlated in terms of the flow variables as below.

$$T_c = 1315.2H^{0.24}/\theta^{1.02}N^{0.88}F^{0.072} \quad [8]$$

The above equation can be simplified after introducing the conditions of the present study (R_i = 27.4 deg, L = 5.9 m, D = 0.147 m) and Saeman's correlation[15] as

$$T_c = 1.2R_iLH^{0.2}/DN\,\theta \quad [9]$$

The above equation considers the dam height in addition to inclination and rotational speed of the kiln and is thus an improvement over the correlation proposed by Saeman.[15]

Unlike the mean residence time, the type of tracer influences the dispersion number. The tracer is represented in terms of its angle of repose and density, consistent with the representation of Chatterjee *et al.*[2] The dispersion number is independent of feed rate of solids and dam height but decreases with increase in rotational speed and inclination of the kiln. The dispersion number is correlated in terms of process variables by the following correlation:

$$(D/UL) = 0.000562R_t^{0.79}/\theta^{0.67}N^{1.06}\rho_t^{0.25} \quad [10]$$

The dispersion coefficients which give a measure of the extent of longitudinal backmixing were calculated using dispersion number and mean flow velocity of solids (ratio of kiln length to the tracer mean residence time) and are listed in Table III. The relations between the dispersion coefficient and feed rate of solids and rotational speed of the kiln are shown in Figures 3 and 4, respectively. It is interesting to note that coal and sand tracers exhibit contrasting behavior with respect to variation of dispersion coefficient, although their mean residence times are similar. The influence of tracer characteristics on dispersion coefficient could be quite complex due to tracer motion in the kiln. The wide variation in the dispersion coefficient of coal can be related to its lower density. While studying the RTD in a horizontal rotary drum reactor, Wes *et al.*[16] have reported an axial dispersion number of 4×10^{-5} m^2/s, which is approximately of the same order of magnitude as that of the present study (2×10^{-5} m^2/s).

B. *Holdup*

Table IV shows the variation in holdup with variation in operating conditions. It can be seen from the table that an increase in slope and rotational speed of the kiln results in decrease in holdup, whereas an increase in feed rate increases the holdup.

Table III. Details of Moment Analysis, Dispersion Number, and Number of Tanks in Series for All of the Runs Performed

Run Number	T_e (min)	T_c (min)	σ (min^2)	σ_θ^2	$(D/UL)_e \times 10^4$	$(D/UL)_c \times 10^4$	D_i, m^2/s $\times 10^5$	n
1	106.6	102.2	7.3	6.43	3.213	3.064	1.75	1556
2	104.1	104.1	5.7	5.28	2.640	3.079	1.47	1894
3	106.1	109.8	6.9	6.10	3.048	3.127	1.67	1640
4	275.3	288.8	145.0	19.13	9.572	10.011	2.02	522
5	326.3	303.1	241.0	22.64	11.333	10.148	2.02	441
6	158.1	156.9	19.5	7.81	3.904	4.804	1.43	1281
7	191.7	194.7	30.3	8.25	4.126	4.566	1.25	1212
8	154.5	137.2	24.8	10.37	5.188	3.624	1.95	964
9	121.8	121.0	9.1	6.12	3.062	2.924	1.46	1633
10	185.3	182.3	29.8	8.67	4.336	4.561	1.36	1153
11	192.1	191.3	34.8	9.43	4.716	4.623	1.42	1060
12	103.4	102.2	8.5	7.91	3.956	3.771	2.22	1264
13	102.9	104.1	9.5	8.93	4.468	3.791	2.52	1119
14	103.0	109.8	7.5	7.06	3.533	3.848	1.98	1415
15	290.3	288.8	219.5	26.06	13.045	12.321	2.61	383
16	293.4	303.1	210.5	24.45	12.239	12.490	2.42	409
17	153.7	156.9	25.0	10.60	5.301	5.913	2.00	943
18	188.7	194.7	38.5	10.81	5.407	5.619	1.66	925
19	118.1	121.0	9.2	6.61	3.305	3.599	1.62	1513
20	181.8	182.3	39.8	12.03	6.018	5.613	1.92	831
21	140.5	—	2.9	—	—	—	—	—
22	193.5	—	23.1	—	—	—	—	—
23	191.5	—	30.3	—	—	—	—	—

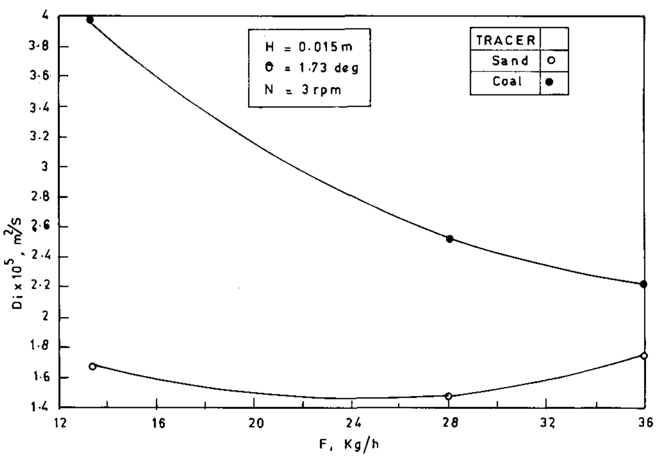

Fig. 3—Typical variation in dispersion coefficient with feed rate of solids.

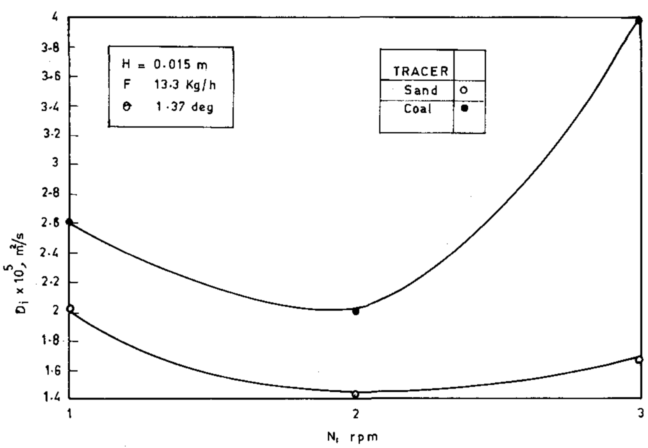

Fig. 4—Typical variation in dispersion number with rotational speed of the kiln.

The variation of holdup with operating variables is represented by the following correlation:

$$W = 66.7F^{0.86}H^{0.4}/\theta^{1.11}N^{0.9} \qquad [11]$$

C. *Material Flow Studies*

As the rotary kiln turns on its axis, the solids bed inside is subjected to transverse motion. The bed motion may be slipping, slumping, rolling, cascading, cataracting, and centrifuging depending on variables such as rotational speed, kiln diameter, percentage fill, bed/wall friction, and bed particle characteristics.[8] In the present study, the bed motion at steady state was either transition between slumping and rolling or rolling. The solids motion in the kiln at 3 rpm was characterized by the rolling regime. On the other hand, when the kiln rotational speed was between 1 and 2 rpm, it was in transition between rolling and slumping.

1. *Effect of bed profile, slope, and rotational speed of kiln on throughput of solids*

Any kiln will have a particular "carrying capacity," depending on the operating conditions. While carrying capacity is fixed independently by rotational speed and inclination of the kiln, the feed rate of solids is an operational variable which can be equal to, greater than, or less than the carrying capacity. When the feed rate is equal to the kiln carrying capacity, the bed profile will be flat; *i.e.*, bed depth will be the same at all points along the length of the kiln. When the feed rate is more than the kiln carrying capacity, the bed profile slopes from the feed end to the exit end, thereby increasing the net inclination of the solids (kiln slope + bed slope). This results in increase in throughput of solids. Similarly, when the feed rate is less than the kiln carrying capacity, the bed will have a negative slope, thereby the net inclination of the solids is decreased, and hence, the throughput of solids also decreases. Thus, the bed height profile is a function of the difference between the feed rate and carrying capacity.

Figure 5 shows a typical variation in bed depth with distance from exit end when the feed rate is less than (line d), greater than (line a), and equal to kiln carrying capacity (lines c and b). If the feed rate is too high, the kiln becomes "flooded," and if it is too low, the kiln will be "starved" of solids.

The variation of throughput of solids with rotational speed, net inclination, and dam height is given by the following relation:

$$F = 6.53 \times 10^4 N^{0.97} (\tan \phi')^{2.17} H^{0.31} \qquad [12]$$

The above correlation is valid when the kiln is under steady-state conditions. Thus, by knowing the inclination, rotational speed of the kiln, and dam height, the

Table IV. Details of Steady-State Bed Profile and Holdup

Run Number	Bed Depth* at (mm) X1	X2	X3	X4	X5	ϕ (deg)	ϕ' (deg)	F (kg/h) Exp.	F (kg/h) Cal.	W (kg) Exp.	W (kg) Cal.
1	35	40	43	48	50	0.615	1.985	36.0	35.1	78.6	70.4
2	31	35	37	38	40	0.382	1.752	28.0	26.8	51.6	56.7
3	22	23	23	23	23	0.042	1.412	13.3	16.7	27.6	29.9
4	33	38	43	45	48	0.615	1.985	13.3	12.1	81.3	80.8
5	26	28	30	31	31	0.212	1.582	6.8	7.4	44.0	45.4
6	27	29	30	31	31	0.170	1.540	13.3	13.6	44.4	43.2
7	28	30	32	35	36	0.340	1.120	13.3	10.1	53.8	56.1
8	25	26	27	30	30	0.233	1.333	13.3	14.8	42.6	38.3
9	40	41	43	45	46	0.255	1.625	28.0	27.5	69.6	72.8
10	36	36	36	36	36	0.000	1.370	13.3	12.8	55.6	55.4
11	29	28	25	24	21	−0.318	1.052	6.8	7.2	32.5	31.2

*X1 = 400; X2 = 650; X3 = 1000; X4 = 1400; X5 = 1750.

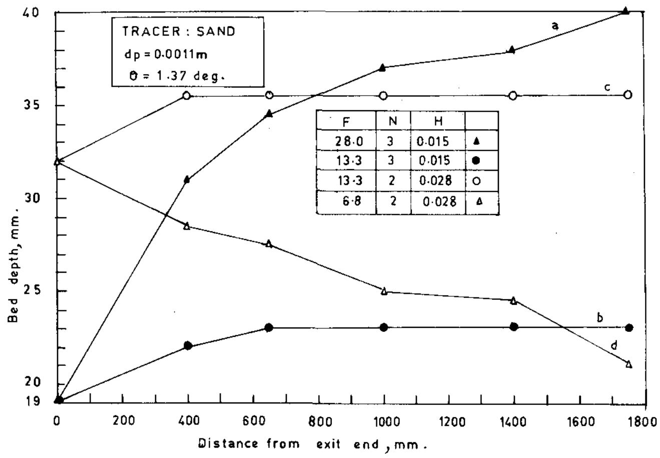

Fig. 5—Variation of bed depth with distance from exit end.

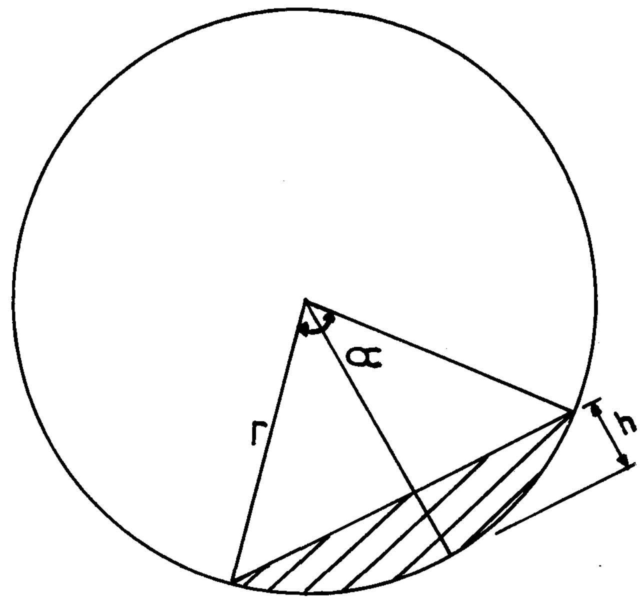

Fig. 6—Schematic diagram for dam height estimation.

feed rate for which the profile will be flat can be computed. The throughput thus computed is compared with the experimental values in Table IV.

During unsteady-state operation, it was observed that the throughput of solids at any point is characterized by a particular and unique profile up to a certain distance from the exit end and was found to be 20 to 25 pct of kiln length, *i.e.*, 1000 to 1200 mm from the exit end of the kiln in the present study. This determines the throughput of solids at any point of time. The bed profile at the remaining length of the kiln does not play much of a role in the throughput of solids at any instant.

2. *Estimation of dam height for a flat profile of solids*

The angle subtended by the bed at the center of the circle (α) is estimated by equating the holdup with the weight of the solids in the bed, as shown by the following equation (Figure 6):

$$W = \rho_i L r^2(\alpha - \sin \alpha)/2 \qquad [13]$$

Then, the dam height can be calculated by

$$H = r[1 - \cos(\alpha/2)] \qquad [14]$$

Thus, by knowing the holdup value, the dam height, which gives rise to a flat profile of solids bed along the length of the reactor for a particular feed rate of solids, inclination, and rotational speed of the kiln, can be estimated. This was tested for the conditions of run 10 (Appendix).

The mean relative quadratic error, SD, between the predicted (X_c) and the experimental (X_e) parameter values defined as

$$SD = \sqrt{\sum_{i=1}^{K} [(X_{c,i} - X_{e,i})/X_{e,i}]^2/(K-1)}$$

where K is the total number of data points, was estimated for all of the equations proposed in the present study; these values are listed below:

Eq. No.	[8]	[10]	[11]	[12]
SD	0.04	0.11	0.06	0.12

IV. CONCLUSIONS

The effect of operating parameters, namely, feed rate of solids, slope and rotational speed of the kiln, type and size of the tracer, and dam height, on the mean residence time, dispersion number, and holdup has been investigated.

1. The operating variables have been correlated as follows:
   a. Mean residence time
   $$T_c = 1315.2H^{0.24}/\theta^{1.02}N^{0.88}F^{0.072} \qquad [9]$$
   b. Dispersion number
   $$(D/UL) = 0.000562R_t^{0.79}/\theta^{0.67}N^{1.06}\rho_t^{0.25} \qquad [10]$$
   c. Holdup
   $$W = 66.7F^{0.86}H^{0.4}/\theta^{1.11}N^{0.9} \qquad [11]$$
2. The mean residence time increases with increase in dam height but decreases with increase in either the rotational speed or the inclination of the kiln.
3. Even though the mean residence time is independent of the type of tracer, the dispersion number is strongly dependent on it. Thus, the spread of distribution is more for coal than sand.
4. The feed rate at which the bed depth profile can be maintained almost flat for a particular rotational speed and inclination of the kiln can be estimated by the following correlation:
   $$F = 6.53 \times 10^4 N^{0.97} (\tan \phi')^{2.17} H^{0.31} \qquad [12]$$

APPENDIX

For Run = 10, W = 55.6 kg (from Table IV). From Eq. [13], $\alpha - \sin \alpha = 2W/\rho_i L r^2 = 0.8307$.

By trial and error, $\alpha = 1.804$. From Eq. [14], $H = r[1 - \cos(\alpha/2)] = 0.028$.

When a dam height of 0.028 m was used, it gave a flat profile of the solids bed ($\phi = 0$).

LIST OF SYMBOLS

D	diameter of the kiln, m
D_i	dispersion coefficient, m^2/s
d_p	average particle diameter, m
D/UL	vessel dispersion number
$E(t)$	RTD function
F	feed rate of solids, kg/h
H	dam height, m
h	bed depth, m
L	length of the kiln, m
N	rotational speed of the kiln, rpm
n	number of perfect mixers in series
R	angle of repose, deg
r	radius of the kiln, m
T	mean residence time, min
t	time, min
W	holdup of solids, kg
α	angle subtended by bed at the center of kiln, radians
θ	inclination of the kiln, deg
ρ	density, kg/m^3
σ^2	variance of a distribution, min^2
σ_θ^2	dimensionless variance of a distribution
ϕ	slope of the bed material, deg
ϕ'	total slope of the bed, $\theta \pm \phi$, deg

Subscripts

c	predicted
e	experimental
i	ilmenite
t	tracer

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