AERODYNAMICS OF MOVING AND STATIONARY BEDS AT HIGH GAS VELOCITIES

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Some features of the operation of a dense bed at high gas velocities are examined, together with the conditions of suspension of the charge in a furnace. The maximum permissible gas velocities in the bed are obtained. The flow friction is determined for loose material in the moving and stationary beds.

The aerodynamics of a dense bed at high gas velocities have an important influence on the operation of shaft and blast furnaces, certain combustion chambers, gas generators, and various drying equipment.

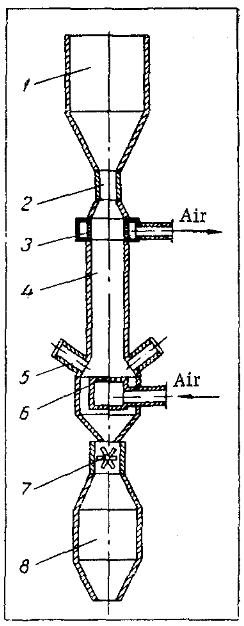

Fig. 1. Schematic of cold counterflow shaft furnace.

However, at sufficiently high gas velocities the bed of loose material ceases to descend and becomes suspended in the shaft. Below, we present the results of experiments to determine the gas velocities in the bed that ensure maximum intensification of the process and normal descent of the material.

The experiments were performed on a cold model of a counterflow shaft furnace (Fig. 1).

The material was periodically loaded into a hopper 1, whence it passed through throat 2 into shaft 4 and encountered an ascending flow of air supplied to the model through a gas distributor 6. The air was exhausted from the model through a mesh 3 with openings 3 mm in diameter. The material was continuously discharged from the model by a feeder 7 into hopper 8. In order to observe the state of the bed above the distributor 6 we installed inspection windows 5. The height of the model shaft could be varied from 350 to 3050 mm, and the diameter from 75 to 120 mm by inserting interchangeable cylinders of different heights and diameters. So as to be able to observe the state of the bed in the model shaft we selected cylinders of transparent plexiglas.

In the experiments the models were loaded in different ways: in the form of "rain," from a hopper, through a cone, etc. The loading method was not observed to have a significant influence on the results.

The static pressures were measured with micromanometers. The air flowrate was measured by means of double orifices.

The porosity of the moving bed was determined by weighing a fixed volume of the moving bed in the shaft of the transparent model. The characteristics of the loose materials investigated are presented in Table 1.

Altogether we conducted more than one hundred experiments. Each set of conditions was repeated 4–5 times. The agreement of the results fell in the range 5–6%.

Visual observations of the moving bed showed that for a gradual increase in the air velocity W_0 in a bed of peas from 0 to 0.98 m/sec over the free cross section of the shaft, the peas descend normally without appreciable changes in the structure of the bed.

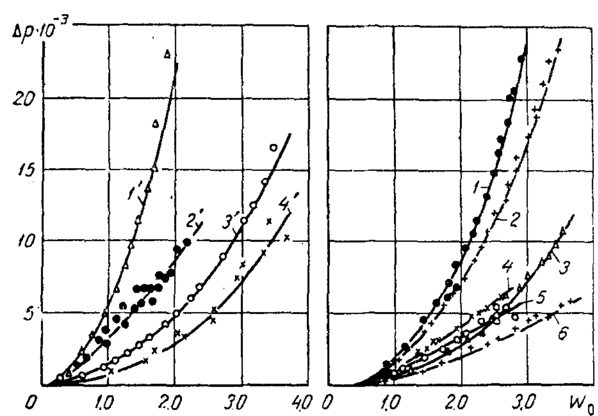

Fig. 2. Pressure loss Δ_p per 1 m of bed thickness (N/m^2) as a function of gas velocity W_0 for the free section (m/sec): 1' and 2') for peas in the stationary and the moving bed, respectively, 3' and 4') the same for pelletized iron ore 4–15 mm in size, 1 and 5) for Agloporit 8–12 mm, 2 and 4) Agloporit 4–20 mm, 3 and 6) Agloporit 12–20 mm.

At a velocity of 0.98 m/sec individual particles start to "boil" (the individual peas tend to shoot upwards). A further increase in velocity leads to an increase in the number of "boiling" peas, and at a velocity of 1.10 m/sec the entire bed of peas is observed to

Table 1

Characteristics and Suspension Velocity of Loose Materials

Material	Apparent specific weight γ_p, kg/m^3	Weighted mean diameter d_M mm	Ratio of shaft diameter to particle diameter D/d_M	Ratio of shaft height to particle diameter H/d_M	Porosity of stationary bed	Theoretical suspension velocity W_s^t, m/sec	Suspension velocity W_s, m/sec	Settling velocity W_l, m/sec	Suspension coefficient K_s	Settling coefficient K_l
Pelletized iron ore ϕ 4–15 mm,	3640	10.56	11.4	33.0—289	0.497	31.20	3.90	3.85	0.125	0.123
Peas Φ 4.6–6.0 mm	1386	5.03	23.9	70—607	0.415	13.35	1.65	1.50	0.124	0.123
Agloporit (spherical cinders) Φ 4–20 mm	1753	11.80	10.2	30—259	0.517	23.10	2.55	2.10	0.110	0.091
Agloporit (spherical cinders) Φ 8–12 mm	1949	9.55	12.6	37—319	0.549	21.80	3.30	2.45	0.151	0.112
Agloporit (spherical cinders) ϕ 12–20 mm	1746	15.50	7.8	23—197	0.553	26.20	3.55	3.0	0.136	0.114
Agloporit (spherical cinders) Φ 4–8 mm	1987	5.45	22.0	64—560	0.583					
Steel balls Φ 12.6–12.9 mm	7778	12.7	9.5	28—240	0.428					
Silica gel Φ 4.5 mm	777	4.73	25.4	74—644	0.329					

Table 2

Effect of Motion of Material on Swelling of Bed

Porosity of material					
steel balls, 12.6–12.9 mm	peas, 4.6–6.0 mm	gravel, 4–6 mm	gravel, 6–10 mm	coke, 6–10 mm	chamotte pellets, 6–13 mm
Stationary bed					
0.428	0.415	0.470	0.495	0.441	0.450
Moving bed (rate of descent 20–120 cm/min)					
0.487	0.473	0.533	0.526	0.541	0.473
0.470	0.445	0.522	0.503	0.556	0.463
0.494	0.475	0.518	0.503	0.530	0.454
0.485	0.465	0.511	0.528	0.541	0.485
0.496		0.504	0.519		0.484
0.482		0.525	0.505		

boil. This process does not appear to complicate the descent of the material (the boiling particles are normally discharged from the model). At a velocity of 1.15 m/sec individual centers of swelling appear due to the intense boiling. This swelling of the bed gradually increases and at 1.20 m/sec becomes massive in character (intense mixing of the material in the lower part of the model is observed). A velocity of 1.24 m/sec marks the beginning of the piston regime, which gradually becomes more and more intense. At a velocity of 1.65 m/sec the bed is suspended at the top of the model (in the region of the head). A further increase in gas velocity leads to the squeezing up (suspension) of the entire column of material. The particles located below the distributor run into the hopper, and those above the distributor form a stationary dense column of material compressed by the air flow. The gas velocity at which the bed is suspended will henceforth be called the suspension velocity W_s.

As the gas velocity is gradually reduced, normal descent of the peas recommences only at 1.50 m/sec, i.e., at a velocity less than the suspension velocity. This velocity will henceforth be called the settling velocity W_l.

A similar picture is observed for beds of other loose materials.

The experiments showed that the suspension velocity calculated for the free cross section of the shaft is more than 8 times smaller than the suspension velocity under free conditions (see Table 1).

The experiments were conducted at $Re_b > 500$, i.e., in the self-similar region.

The theoretical suspension velocity (critical velocity for a free individual particle) was calculated from the equation

$$W_s^t = \sqrt{\frac{4g}{3}\frac{\gamma_p}{f_p}\frac{d_м}{\gamma}} = 3.62\sqrt{\frac{\gamma_p}{f_p}\frac{d_м}{\gamma}}. \qquad (1)$$

In the experiments we investigated loose materials of approximately spherical shape ($f_p = 0.43$). For nonspherical loose materials f_p can be taken, for example, in accordance with Vakhrushev's equation [6].

Table 1 shows that W_s slightly exceeds the velocity of onset of fluidization. Thus, for example, for the peas tested in the experiments the velocity of onset of fluidization is 1.1–1.2 m/sec, while the suspension velocity is 1.65 m/sec. The "clear section" for the passage of gases in the bed is probably very small, and close to the edge of the suspended bed, a velocity field is formed that considerably exceeds the velocity calculated for the free cross section of the shaft.

The suspension coefficient K_s, determined experimentally (Table 1), enables us to pass from the theoretical to the real suspension velocity

$$W_s = K_s \cdot 3.62\sqrt{\gamma_p d_м / f_p \gamma}. \qquad (2)$$

Correspondingly, the settling velocity

$$W_l = K_l \cdot 3.62\sqrt{\gamma_p d_м / f_p \gamma}. \qquad (3)$$

The maximum permissible gas velocity in the bed is apparently a gas velocity close to the piston regime. As the experiments showed, the descent of material then proceeds without any difficulty, and the gas velocity in the bed is quite high.

The experiments established that this velocity is

$$W_{opt} = 0.75W_s = K_s \cdot 2.72\sqrt{\gamma_p d_м / f_p \gamma}. \qquad (4)$$

It was experimentally established that W_s, W_l, and W_{opt} do not depend on the heights and diameter of the apparatus.

We also investigated the flow friction for different materials in both stationary and moving beds at ordinary and elevated gas velocities (Fig. 2). The experimental data for the stationary bed are in good agreement with results of other investigators [1–5]. The figure shows that the quantity Δp is to a considerable extent determined by the fractional composition of the material: for fine materials (peas) it is considerably higher than for coarse ones. Moreover, for all the materials tested discontinuities of the Δp-W_0 curves were not observed for free and suspended stationary beds. The above-mentioned law is particularly well expressed by the Δp-W_0 curve for peas, whose suspension velocity, equal to 1.65 m/sec, was exceeded by more than a factor of 2. It follows from these data that the structure of the suspended bed (its porosity, the orientation of the particles) is identical in character with the free stationary bed.

The gas pressure losses in the stationary bed are described by the following equation:

$$\Delta p = \xi \frac{\gamma W_e^2}{2gd_e} l. \qquad (5)$$

As the experiments showed, the resistance of the moving bed is considerably less than that of the stationary bed due to the increase in porosity (Table 2) and the orientation of the particles for minimum resistance (decrease in mid-section). These data coincide with results of several other investigators [7–9].

We shall introduce into Eq. (5) the corresponding correction factors and obtain the equation for calculating Δp in a moving bed of loose material,

$$\Delta p = K_{or} K_{por} \xi \frac{\gamma W_e^2}{2gd_e} l, \qquad (6)$$

where the quantities W_e and d_e can be taken from the data for a stationary bed, while the orientation coefficient K_{or} and the porosity coefficient K_{por}, which depend on numerous factors (particle size, shape, etc.), must be experimentally determined in each individual case by blowing through the stationary and moving beds of loose material.

For the material tested the product K_{or} K_{por} is: peas 0.57, for pelletized iron ore 0.75, for Agloporit 8–12 mm in diameter 0.38, and for Agloporit 4–20 mm in diameter 0.64. In fact, as industrial practice has shown, calculation of the pressure loss in the bed from the formulas for a stationary bed does not coincide with the real values of the resistance of the moving bed measured in functioning apparatus. What has been said is illustrated by Fig. 2, from which it is clear that the difference in Δp for moving and stationary beds of certain loose materials approaches 50% and more.

A study of the distribution of static pressures over the height of the shaft showed that the dependence of Δp on the thickness of the bed has a linear character for both moving and stationary beds.

SUMMARY

1. We have experimentally established the maximum permissible gas velocities in a bed, together with the conditions of suspension of the charge in a furnace.

2. It has been established that the gas velocity at which the bed is suspended is approximately 1.5 times greater than the velocity of onset of fluidization, which exceeds by many times the actual working gas velocity in industrial apparatus and makes it possible to increase by several times the productivity of existing equipment.

3. It has been established that the flow friction in the moving bed is much less than in the stationary bed, which is attributable to the orientation of particles for minimum resistance and the increase in the porosity of the bed when it is in motion.

4. Calculation of the flow friction for a moving bed from formulas and drag coefficients obtained for a stationary bed of loose material is inaccurate. The error may reach 50% and more.

5. Motion of the bed leads to an increase in its porosity even in the absence of a gas flow through the bed.

NOTATION

γ_p—apparent specific weight of material, particle, kg/m^3; d_M—weighted mean diameter of particle, m; f_p—particle drag coefficient; γ—specific weight of gas flow, kg/m^3; ξ—drag coefficient of bed according to N. M. Zhavoronkov [1]; $W_e = W_0/\varphi$—equivalent gas velocity in bed; $d_e = \sqrt{F/0.785}$—equivalent particle diameter; l—height of bed; φ—porosity of bed; F—mid-section of particle.

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