FLUIDIZED BED WITH BIG ANGLE OF INCLINATION OF A GAS DISTRIBUTOR

by

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A multi-sectional through-flow apparatus, proposed in [1], is described. In this apparatus stable fluidization of particles and deep and uniform cooling or thermal treatment is ensured by use of an inclined gas distributor featuring an almost zero pressure drop. The pattern of particle circulation was investigated. Formulas for calculating of fluidization velocity at the start of particle movement and for computing the pressure drop and rate of particle cooling by gas were derived. Characteristics of operating commercial apparatus are given.

Introduction

The heat treatment of large particles with high internal thermal resistance is most efficient in a multisectional apparatus. Historically uniform fluidization of the material in adjacent sections receiving gas from the same blower can be produced only at a very high pressure drop across the gas distributor. Energy consumption for blowing such a design is prohibitive.

Through-flow fluidized bed

The problem of the heat treatment of large particles can be solved if fluidization is effected on an inclined gas distributor arranged at an angle close to the angle of repose of the material, Fig. 1a. The apparatus consists of sections with vertical partitions, leaving a gap between each partition and the distributor. The discharge is outfitted with a bridge which controls the material layer thickness on the distributor. In each section formed by two partitions and a portion of the inclined distributor, the upper level of the fluidized bed is almost horizontal, *i.e.* the bed height is different at opposite partitions. Owing to this height difference each section has a zone of active spouting, where particles are brought from the gas distributor to the bed top, and a zone of quiet fluidization, where particles sink slowly to the gap under the partition. The inclined position of the distributor adds to the apparatus stability with respect to random fluctuations in the bed height in any of the sections. For example, when the height is increased in the last (discharge)

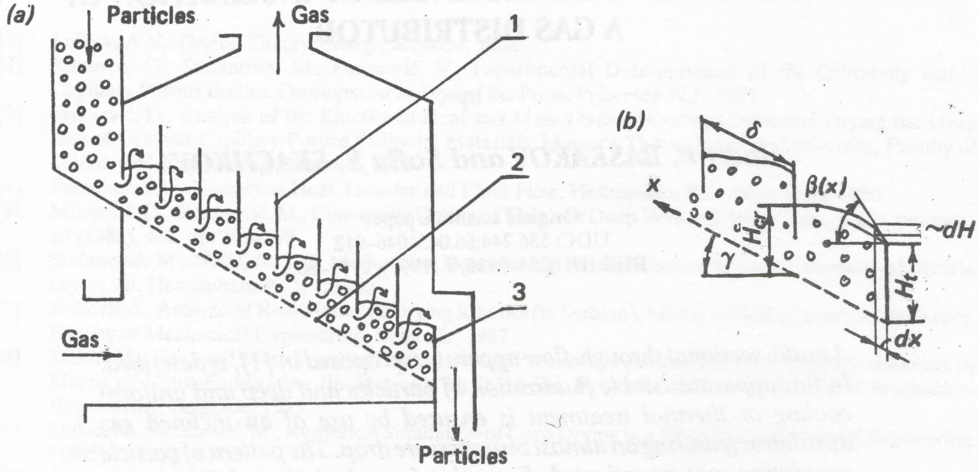

Figure 1. Scheme of apparatus (a) and detalization of construction (b)
1 – partition, 2 – gas distributor, 3 – bridge

section as a result of changing the bridge height, more gas flows to the upstream sections, where the bed height has not yet changed. Flow of the material through the gaps in those sections is enhanced and the sections are filled rapidly to the height that corresponds to the new position of the bridge.

Gas velocity and bed pressure drop at the start of particle movement

To determine the pressure under the distributor and the gas flow rate required to fluidize the bed, consider the profile of the bed upper boundary in the last section (Fig. 1b). When the bed is at rest, the profile is determined by the angle of repose β, which, according to [2], decreases as the rate of gas filtration through the dense bed is raised. Studies made on iron-ore pellets shows that the dependence of tgβ on the ratio $W = U/U_{mf}$ approaches parabolic

$$\mathrm{tg}\beta = (1 - W^2)\,\mathrm{tg}\beta_0 \qquad (1)$$

Here: β_0 is the angle of repose without gas filtration; U_{mf} is an incipient fluidization velocity in the apparatus with a horizontal gas distributor. For the calculation of U_{mf} see [3].

In every section gas velocity changes along the gas distributor owing to variations in the bed height; as a result β depends on x. Proceeding from geometric considerations, a change dH of height on length dx is

$$dH = [\mathrm{tg}\beta(x) - \mathrm{tg}\gamma]\cos\gamma\, dx \tag{2}$$

The bed height is the lowest at the left-hand partition of the discharge section (see Fig. 1b). If the pressure under the distributor increases, the maximum gas flow occurs where the bed height is a minimum. At this location bed fluidization first occurs, $U = U_{mf}$ and surface becomes horizontal, $\beta = 0$.

If the gas flow rate increases further, particles will be thrown up where the bed depth is the least and will roll over the bed and drop over the bridge. The discharged material will be replaced by material flowing from the adjacent upstream section through the gap between the partition and the distributor. When the gas velocity through the layer of pellets is about 5 m/s the Reynolds number, Re, exceeds 4000; therefore simplifying the Ergun [4] formula, one obtains

$$U = \{p\varepsilon^3 d/[1.75\rho_g H(1-\varepsilon)]\}^{0.5} \tag{3}$$

Since $U \sim H^{-0.5}$, disregarding the flow of the gas parallel to the distributor in the section, expression (1) can be rewritten as

$$\mathrm{tg}\,\beta(x) = \mathrm{tg}\beta_0(1 - H_{cr}/H) \tag{4}$$

where H_{cr} is the bed height at the partition when the material starts moving.

Substition of (4) into (2) yields

$$dH = [\mathrm{tg}\beta_0(1 - H_{cr}/H - \mathrm{tg}\gamma)]\cos\gamma\, \mathrm{d}x \tag{5}$$

Integrating (5) with a boundary condition ($x = 0$, $H = H_t$), one derives an equation for computing the distribution of the bed height over the section length x at the moment particle motion starts

$$\ln[(H/H_{cr} - Z)/(H_t/H_{cr} - Z)] = (x\mathrm{tg}\beta_0\cos\gamma/ZH_{cr} + (H_t - H)/H_{cr})/Z \tag{6}$$

which, on substituting $x = \delta$, makes it possible to find H_{cr}. Here $Z = 1/(1 - \mathrm{tg}\gamma/\mathrm{tg}\beta_0)$.

The pressure required to fluidize a column of material of height H_{cr}, *i.e.*, corresponding to the onset of material movement is

$$p_{mf} = \rho_p(1 - \varepsilon)H_{cr} \tag{7}$$

Knowing the distribution of the bed height over the section length, formula (3) can be used to calculate the gas velocity distribution at the mean filtration rate required for the material to start moving.

When the distributor inclination angle γ is equal to the angle of repose β_0, the bed height H_{cr} at $x = \delta$ is found from (6)

$$H_{cr} = \delta \sin\gamma\left[\sqrt{1 + (H_t/\delta\sin\gamma)^2} - 1\right] \tag{8}$$

The distribution of the gas filtration velocity over the section length is

$$U_{bm} = U_{mf}\left[(H_t/H_{cr})^2 - 2x\sin\gamma/H_{cr}\right]^{-0.25} \tag{9}$$

The length average gas velocity at the moment material starts moving is

$$\overline{U}_{bm} = 4U_{mf}\left[(H_t/H_{cr})^2 - 1\right]^{-1}\left[(H_t/H_{cr})^{3/2} - 1\right]/3 \tag{10}$$

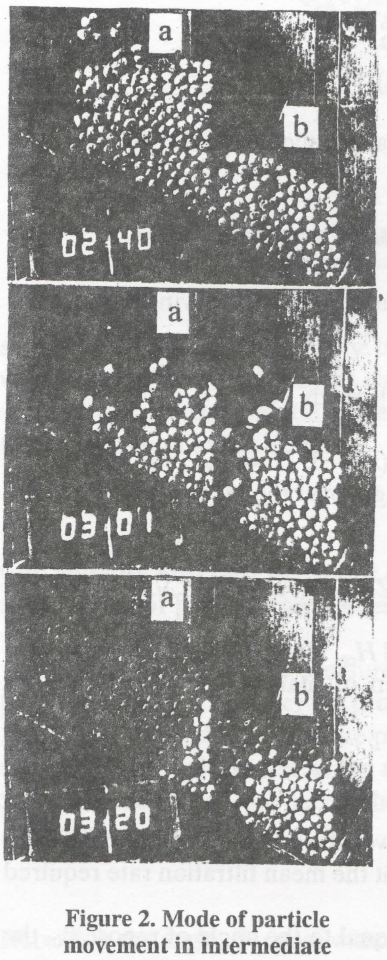

Figure 2. Mode of particle movement in intermediate (a) and last (b) cells
$d = 12$ mm, $\rho_p = 2970$ kg/m^3; $H_t = 0.25$ m; $\sigma = 0.3$ m; $\gamma = 25°$; $\overline{U} = 7.8$ m/s; $p = 4.25$ kPa

Particle movement pattern

As the mean gas velocity is raised above $\overline{U}_{bm}$, the number of particles moving from one chamber to another and over the bridge in the discharge chamberi, *i.e.* the apparatus output grows. Experimetns show that the dependence of output on the flow rate is close to linear. This feature facilitates control and allows maintaining the temperature conditions approximately constant at various flow rates of particles cooled or heated in the apparatus.

Particle movement was studied in a laboratory model 0.1 m wide by filming (24 frames per second) through a glass wall (Fig. 2). Particle trajectories drawn from films are shown in Fig. 3a and 3b. In intermediate chambers, particles rise in the zone of active fluidization at a speed of 1.5 to 3.5 m/s and sink at the opposite partition at 0.001 to 0.050 m/s. Part of the sinking material is again drawn into the ascending flow, thus forming a circulation circuit. The rest of the material passes through the gap into the next section. In the final (discharge) section, particles rise at a speed varying from 1.2 m/s at the partition to 0.05–0.8 m/s along the stagnation zone boundary (shaded in Fig. 3b).

A formula was derived experimentally to calculate output in kg/s of the apparatus for cooling of iron-ore pellets:

$$G = b(29\sigma - 52H_t + 0.15\gamma + 28W_m - 40) \tag{11}$$

where b is the distributor width, m; W_m is the modified fluidization index equal to the ratio between

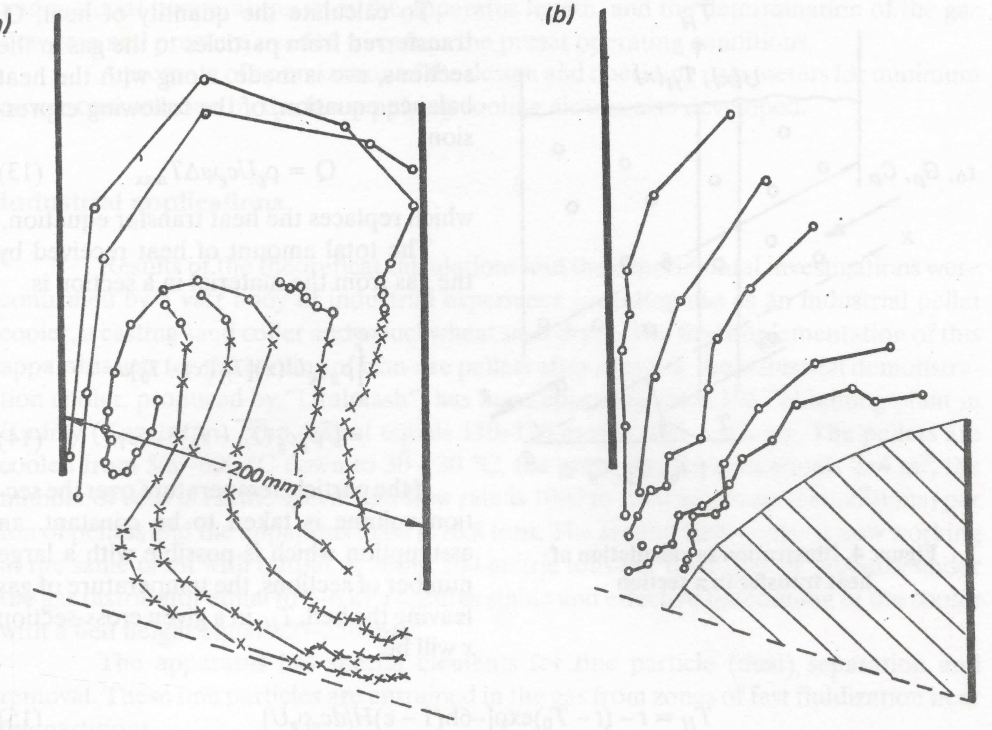

Figure 3. Movement trajectories of pellets in intermediate (a) and discharge (b) sections of the model. The distance between dots is the path traveled by a pellet in 1/24 s, and between crosses is that traveled in 1 s. $\overline{U}$ = 8.66 m/s

the mean gas flow velocity and the velocity at which pellets start moving in the apparatus. As the gas flow rate is raised, the pressure drop across the bed on the inclined distributor exceeds the pressure p_{mf} corresponding to the onset of the particle movement. Experimental data yields

$$p = p_{mf} + 3\beta + 1.3W_m - 1.8 \tag{12}$$

Heat transfer between the gas and particles

Since the gas velocity is distributed nonuniformly over the length of each section, the particle-to-gas heat transfer coefficient varies also. Under these conditions it is more convenient to use the heat exchanger efficiency $\psi = (\overline{T_H} - T_0)/\Delta T_{\max}$ [5] characterizing the degree to which the heat-transfer agent potential is utilized. Here T_0 and $\overline{T_H}$ stand for the gas temperature under the bed and at the section outlet (mean), while $\Delta T_{\max} = t - T_0$, where t is the particle temperature at the section outlet (Fig. 4).

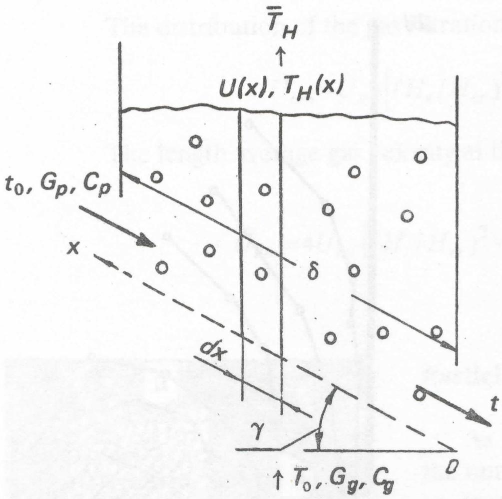

Figure 4. Illustration to calculation of heat transfer in a section

To calculate the quantity of heat, Q, transferred from particles to the gas in the sections, use is made, along with the heat balance equation, of the following expression:

$$Q = \rho_g U c_g \psi \Delta T_{max} \tag{13}$$

which replaces the heat transfer equation.

The total amount of heat received by the gas from the material in a section is

$$\int_0^{\delta\cos\gamma} \rho_g c_g U(x)[T_H(x) - T_0]dx = \overline{U}\rho_g c_g (\overline{T_H} - T_0)\delta\cos\gamma \tag{14}$$

If the particle temperature over the section volume is taken to be constant, an assumption which is possible with a large number of sections, the temperature of gas leaving the bed, T_H, in a given cross-section x will be

$$T_H = t - (t - T_0)\exp[-6h(1-\varepsilon)H/dc_g\rho_g U] \tag{15}$$

where h is the particle-to-gas heat transfer coefficient. Substituting T_H into expression (14) and re-arranging it, one has

$$\psi = (\overline{T_H} - T_0)/(t - T_0) = 1 - (\delta\cos\gamma)^{-1}\int_0^{\delta\cos\gamma} U\overline{U}^{-1}\exp(-6h(1-\varepsilon)H/d\rho_g c_g U)dx \tag{16}$$

The heat transfer completeness ψ can be calculated for given distributions of velocity U, heat transfer coefficient h, bed height H and porosity ε over the section length. It is often simpler to determine ψ directly from experiments.

Proceeding from experiments on cooling of iron pellets in a model 0.1 m width and with δ varying from 0.2 to 0.4 m, γ from 14 to 30°, and W_m from 1.6 to 2.2, we derived the formula

$$\psi = 1.05 + 0.4\delta - 0.004\gamma - 0.26W_m \tag{17}$$

The above results served as the basis for development of a FORTRAN program intended for the design of industrial coolers. The program allows the determination of the distributor width and the number of sections ensuring the required output of the apparatus and preassigned cooling efficiency, calculation of the distribution of the gas and material temperatures over the apparatus length, and the determination of the gas flow rate and pressure needed to ensure the preset operating conditions.

A program of optimization of the design and operating parameters for minimum power consumption by the fan supplying cooling air was also developed.

Industrial applications

Results of the theoretical calculations and the experimental investigations were confirmed by a vast body of industrial experience including use as an industrial pellet cooler, a casting sand coller and a buckwheat seed dryer. The first implementation of this apparatus was for the cooling of iron-ore pellets after roasting. An industrial demonstration cooler, produced by "Uralmash", has been operating since 1985 at mining plant in Rudniy (Kazahstan). The output equals 110–120 metric tons per hour. The pellets are cooled from 500–600 °C down to 30–130 °C, the grate surface area equals 2×4 m^2, the number of cells is 12, the specific air flow rate is 1000 to 1500 m^3 (normal conditions) per ton of pellets and the apparatus mass is 16,8 tons. The second such cooler is now working at the same plant with output of about 200 metric tons per hour. An air pressure under the gas distributor equal to 5–6 kPa ensures stable and effective functioning of the cooler with a bed height equal 200–250 mm.

The apparatus has special elements for fine particle (dust) separation and removal. These fine particles are entrained in the gas from zones of fast fluidization near the partitions.

With the same technological characteristics as a packed bed cooler and the same specific electricity consumption for air blowers, this apparatus cools pellets in a 4 to 5 minutes instead of 60 minutes in the packed bed cooler. Also the apparatus mass is an order of magnitude less.

"Uralmash" solved the problem of cooling of casting sand after drying and roasting by using this type of apparatus. A simple cooler with a 40 ton per hour output substituted for four coolers having ordinary fluidized beds. The main drawbacks of the ordinary fluidized bed coolers were uneven fluidization with stagnant zones, noise produced by high-speed conveyers transporting sand to the bed, and dusting. In comparison this apparatus, having no moving parts, is hermetically sealed and produces no noise *at all*. Also, in the process of transport of sand from cell to cell the sand is cleaned of clay particles, which are entrained by air into a separation system. This cleaning improves the sand quality and diminishes dusting in the process of transport after the cooler.

A similar apparatus is used for drying of buckwheat seeds after thermal treatment by steam and before treatment in a special installation to remove husks. In this case the apparatus permits combining drying of buckwheat with cleaning from sweepings and fine admixtures.

The apparatus has an output of 600 kg per hour, lowers the buckwheat water content from 16–18% down to 12–14%, and uses a drying agent (air) with a temperature of 150 to 180 °C, the output buckwheat temperature being no more than 50 °C.

A dryer for barley seeds with a 3 ton per hour output was also constructed. An external view of the bed of barley seeds in the laboratory installation is shown in Fig. 5. The seeds have elongated form, high angle of repose and a tendency to produce stagnant zones. However, if the angle of inclination of the gas distributor equals to 35° and height of the gaps equals 120 mm, steady motion of seeds along the gas distributor is achieved, an active stirring being produced in the process of seed flow from cell to cell.

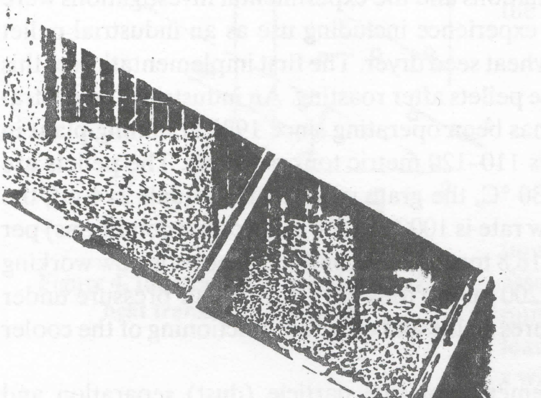

Figure 5. Movement of barley seeds in a laboratory model

Conclusion

A multi-sectional apparatus with a fluidized bed on an inclined gas distributor can be used for gas heat-and-mass treatment of a wide class of dispersed material. Such an apparatus has many advantages in comparison with ordinary fluidized bed installations including the following:

- produces deep and uniform treatment because of the same dwell time of all particles in the active zone,
- uses an advantageous cross-flow scheme of a heat transfer,
- has low blower power consumption because of low gas distributor pressure drop,
- has no moving parts so can be easily hermetically sealed,
- has a simple system of automatic control,
- can be easily installed in an existing technological line because of its compactness.

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Nomenclature

b – distributor width
c_g – gas heat capacity
d – particle diameter
G – apparatus output
H – bed height
H_t – bridge height
H_{cr} – bed height at the partition when the material starts moving
h – heat transfer coefficient
p – pressure under the bed
p_{mf} – pressure under the bed at the moment the material starts moving
T – gas temperature
t – particle temperature
U – gas velocity
U_{mf} – minimum fluidization velocity
U_{bm} – gas velocity at the moment the material starts moving
x – coordinate along the distributor

Greek letters

β, β_0 – angle of repose with and without gas filtration
γ – distributor inclination angle
δ – partition spacing
ε – bed porosity
ρ_g, ρ_p – density of the gas and particles

Nondimensional numbers

Re – Reynolds number,*
W – fluidization index,*
W_m – modified fluidization index,*
ψ – heat exchanger efficiency,*

*

$\mathrm{Re} = Ud/\nu$
$W = U/U_{mf}$
$W_m = \overline{U}/\overline{U}_{bm}$
$\psi = (\overline{T_H} - T_0)/(t - T_0)$

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