Design and Operation of Convective Industrial Dryers

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Design and operational performance of convective industrial dryers are an important field of chemical engineering, which is still governed by empiricism. This article addresses the design vs. operation problem for three basic types of continuous convective industrial dryers: conveyor-belt, fluidized bed, and rotary. Design procedures determined the optimal construction and operational characteristics in terms of total annual cost for each type involved and for a given production capacity through appropriate mathematical modeling. All dryer types were compared by evaluating optimum configurations for a wide range of product characteristics and production capacity values. Once the dryer configuration was specified, its operational performance was evaluated by comparing the optimum operational cost vs. production capacity for predefined optimum designed structures. Rotary dryers were more expensive to design than fluidized bed dryers. Operationally, however, it is the other way around due to the favored heat transfer achieved in rotary dryers. Conveyor-belt dryers lie somewhere between producing satisfactory results in terms of both design and operation. Case studies on foods and inorganics are included to demonstrate the performance of each process as well as the effectiveness of the proposed approach.

Introduction

Process design is principally a combination of structural and parametric optimization efforts, carried out under flowsheeting constraints. In the case of dryers, it has become an increasingly challenging problem, since most efforts in the field face problems of extreme difficulty related to the complex description of the numerous interconnected and opposing drying phenomena (Kiranoudis et al., 1995a). Selection of the appropriate dryer type is one of the most complex and poorly understood areas in drying technology. It has been largely neglected in the literature, due to the difficulty encountered in describing it quantitatively.

There are a large number of dryer types available, and within each type there are various options. The choice is rarely obvious; frequently there are several dryers that can dry the material acceptably in different ways. The basic requirement for the dryer is that it must achieve the required amount of drying in an acceptable time, be capable of handling the material, and produce acceptable product quality, with equipment of a suitable size and cost which meets all the relevant safety and environmental requirements. Williams-Gardner (1971) and Van't Land (1984, 1991) used simple-tree methods for dryer selection. Noden (1969) and Papagiannes (1992) have outlined methods based on their experience in industry. However, both procedures are largely qualitative and only cover a few broad types of dryers. Strumillo and Kudra (1986) have made significant advances in the classification of materials, but the resulting algorithms are largely untried in industrial practice. Nevenkin and Gavdarov (1992) have developed a method based on neural network principles, using details of which dryer type has been successful in the past in various applications. Kemp and Bahu (1995) proposed a dryer selection procedure based on a five-step algorithm to analyze the problem that was especially suitable for incorporation into an expert system.

All dryer selection techniques perform an extensive rule-based search in corresponding equipment trees. The resulting rival types have to be compared against quantifying criteria to make definite decisions. Design of dryers is the only way to quantitate the rival solutions and convey selection methodologies in decision-making strategies.

Myklestad (1963) developed a procedure for predicting the material moisture content profile of solids along the length of a rotary dryer by assuming that gas temperature is linearly related to their moisture content. For the same type of dryer, a more fundamental approach used by Sharples et al. (1964) required the simultaneous solution of four differential equations describing heat and mass balances within the dryer. Ahn et al. (1964) used dynamic programming techniques to study optimal air distribution patterns in cross-flow grain dryers. Thompson (1967) developed multivariable search techniques, based on single-dimensional search algorithms, for use in studying the optimal design of convective grain dryers. Davidson et al. (1969) extended the work of Porter (1963) and Turner (1979) to include mass transfer to rotary-dryer models. Thorne (1979) combined the work of Kelly and O'Conell (1977) on retention time with the vapor-diffusion model of Garside et al. (1970) to develop a computer-simulation model on a single-pass, open-center rotary dryer. A simplified rotary-dryer drying model was proposed by Kisakurek (1982) by assuming constant solids temperature and neglecting sensible heat effects. In a computer program developed by Platin et al. (1982) external convective heat transfer to solids was assumed to control the drying process. Kamke and Wilson (1986) developed a mathematical model for simulation purposes for rotary dryers of wood. It included studies on retention time and heat- and mass-transfer phenomena within the dryer.

Farmer (1972) developed a dynamic programming algorithm for the single-staged concurrent flow dryer with a counterflow cooler. The objective function considered energy cost and used grain quality constraints. Thygenson and Grossmann (1970) presented a mathematical model for the modeling and optimization of a through-circulation packed-bed dryer. Brook and Bakker-Arkema (1978) determined the optimum operational parameters and size of two-stage and three-stage concurrent-flow grain dryers with intermediate tempering stages. The objective function was based on energy and capital costs. Becker et al. (1984) used a simple process model, applicable for microcomputer-based on-line applications, to optimize the operation of the dryer. Bertin and Blazquez (1986) presented a mathematical model for a tunnel-dehydrator of the California type for plum drying, and optimized the production rate of the dryer. Kaminski et al. (1989) used two methods of multiobjective optimization in order to analyze the process conditions of L-lysine drying in a fluidized-bed dryer. Results obtained were compared to those of one-objective optimization. Chen (1990) developed a mathematical model based on liquid-diffusion theory and basic heat and mass-transfer principles to simulate and optimize a two-stage drying system that involved a fluidized- and a fixed-bed dryer. Vagenas and Marinos-Kouris (1991) presented a mathematical model for the design and optimization of an industrial dryer for Sultana grapes. The optimal conditions were evaluated by minimizing the thermal load of the dryer per unit mass of dry product. Douglas et al. (1993) developed a mathematical model for the dynamic simulation and design of a rotary dryers applied to sugar crystalline processes. Comparison of model predictions with industrial data produced accurate predictions for steady-state operation and dynamic trends that are consistent with engineering judgment. Jumah and Mujumdar (1993) developed a mathematical model suitable for simulation and preliminary design of a continuous well-mixed fluid bed dryer. Kiranoudis et al. (1994a) developed a mathematical model suitable for the design and optimization of conveyor-belt dryers. The objective of the design was the evaluation of optimum flow-sheet structure, construction characteristics, and operational conditions. The methodology was further expanded to include production planning analysis of multiproduct dehydrating plants (Kiranoudis et al., 1994b) and flexible design under production planning criteria for conveyor-belt dryers (Kiranoudis et al., 1993, 1995b). Maroulis et al. (1995) developed a fluidized-bed dryer simulator, based on a mathematical model describing heat and mass transfer within the dryer.

Among the numerous dryer families existing, the most popular is the one involving convective drying. In this method, the sensitive heat of a gaseous medium is supplied to the material surface by convection. The drying agent flowing past or through the body also removes the evaporated water and transports it from the dryer. As a drying agent, hot air is frequently used, but other media, such as combustion waste gases, can also be applied. Continuous convective industrial dryers include three basic dryer types: conveyor belt, fluidized bed, and rotary. Convective drying applies to virtually all sectors of the chemical industry involving the dehydration of materials ranging from food products to inorganic minerals. This work addresses important design and operational aspects of these three convective industrial dryer types in detail. The process was described by deducing a corresponding mathematical model. The construction and operational performance for a certain level of production capacity and for a wide range of product characteristics are evaluated by appropriately optimizing the total annual cost resulting from the construction of a new plant. Optimum configurations were evaluated in a straightforward way for all dryer types. Their corresponding operational performance was evaluated by comparing the optimum operational cost vs. production capacity for predefined optimum designed configurations.

Mathematical Modeling

An industrial conveyor-belt dryer typically consists of drying sections placed in series. All sections participating in the dryer are equipped with a common conveyor belt, on which the product to be dried is uniformly distributed at the entrance, while redistribution of the product takes place when it leaves a drying section and enters the one following. Each section is equipped with an individual heating utility and fans for air recirculation throughout the product. Combustion gases coming from a fuel burner are typically used for heating air, which on entering the section is mixed with the recirculation. It is common practice that within each section, temperature and recirculation of the air drying stream are controlled. In this case, the final control elements are the fuel flow valve and the section dampers which regulate the burner combustion and the flow rate of air streams within the drying section, respectively. A typical diagram of the interior of a drying section, as well as the arrangement of its control facilities, is presented in Figure 1.

Since each dryer is made up of similar modules, the overall steady-state mathematical model will be formed by repetition of the modules that each individual component contributes. The mathematical model of each drying section involves heat

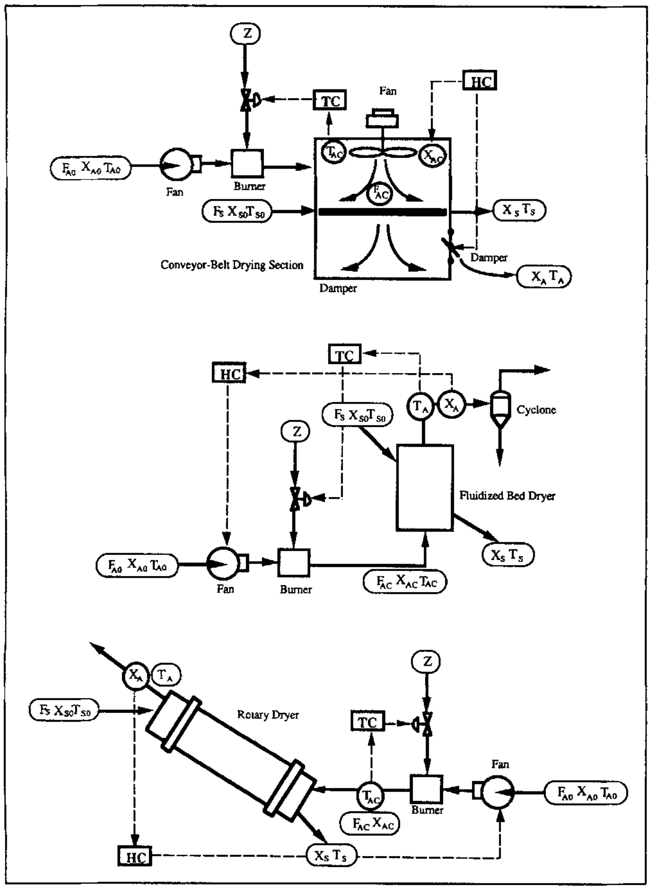

Figure 1. Convective industrial dryers (conveyor-belt drying section, fluidized bed, and rotary).

and mass-transfer balances of air and product streams, as well as heat and mass-transfer phenomena that take place during drying, and is constrained by product quality and equipment construction characteristics. The humidity balance in the drying section and its compartment is expressed by the following relations, respectively:

$$F_{AC}(X_A - X_{AC}) = F_A(X_A - X_{AM}) \tag{1}$$

$$F_{AC}(X_A - X_{AC}) = F_S(X_{S-} - X_{S+}). \tag{2}$$

The corresponding heat balance, assuming negligible heat losses, is expressed as follows:

$$F_{AC}(h_A - h_{AC}) = F_A(h_A - h_{AM}) \tag{3}$$

$$F_{AC}(h_A - h_{AC}) = F_S(h_{S-} - h_{S+}). \tag{4}$$

Heat and mass-transfer phenomena during drying are indeed complicated and their solution demands considerable computational time. They involve coupled-transfer mechanisms both within the solid and the gas phase. In this case, a simplified model is considered. It has an exponential form and contains a mass-transfer coefficient of a phenomenological nature, which is usually called the drying constant. This constant chiefly accounts for mass diffusion within the solid phase, but also embodies boundary-layer phenomena when it is considered to be a function of all process variables affecting drying. Ample accuracy is combined with sufficient low computation time (Kiranoudis et al., 1992a). On the basis of the preceding, mass transfer is expressed by the following equation (Kiranoudis et al., 1992b):

$$X_{S+} = X_{S-} \exp(-k_M t). \tag{5}$$

Heat transfer is chiefly controlled by the heat-transfer coefficient at the air boundary layer. For the purpose of developing the particular mathematical model, it is assumed that the heat-transfer coefficient takes a value high enough to allow the product stream leaving the section to be in thermal equilibrium with the air stream leaving the product. This assumption removes the need for an unnecessary differential equation that would not improve the model significantly. On the basis of the preceding, heat transfer within the section is expressed by means of the following equation:

$$T_A = T_{S+}. \tag{6}$$

The distribution of the product on the conveyor belt is characterized by the belt load variable. This variable is expressed in units of mass of product per unit of belt area. Its value at the entrance of each drying section can be calculated by means of the following equation:

$$\rho_B = \frac{F_S(1 + X_{S+})t}{A} \tag{7}$$

A simplified model of the actual burner was taken into consideration (Maroulis et al., 1995). Assuming that the fuel is a hydrocarbon, the production rate of water by means of the combustion reactions is given by the following equation:

$$R_W = 9C_H z. \tag{8}$$

The overall mass and moisture balance over the burner are given below:

$$F_{A0} + z = F_{AC} \tag{9}$$

$$F_{A0}X_{A0} + R_W = F_{AC}X_{AC}. \tag{10}$$

The corresponding energy balance, assuming negligible heat losses, is given as follows:

$$F_{A0}h_{A0} + z\Delta H_F = F_A h_{AM}. \tag{11}$$

The temperature diminution of the drying air stream when passing through the solid particles is given by the following equation:

$$\Delta T_A = T_{AC} - T_A \tag{12}$$

This diminution should not exceed a maximum value that would guarantee uniform drying, because it prevents creation of axial mass and temperature gradients within the solid particles. The electrical power consumed by the fans is expressed by the following relation:

$$E = \Delta P F_{AC}. \tag{13}$$

The recirculation of the drying air stream is given as

$$R = F_{AC}/F_A. \tag{14}$$

The economic evaluation of the dryer is based on the determination of its total annual cost. The corresponding capital cost is affected by the conveyor-belt area, the capacity of the burner in terms of fuel flow rate, and the installed power of fans involved. Furthermore, the capital cost is affected by the cost of the conveyor belt, which is determined by its corresponding total area within each drying section examined. All capital cost components obey economy-of-scale laws, that is to say, expansion of the unit size with respect to its characteristic dimensions will contribute reduced additional capital cost, nonproportional to the actual size expansion:

$$C_{CP} = \alpha_{D,CV}\left(\sum_{\text{Sections}} A\right)^{n_{D,CV}} + \sum_{\text{Sections}} \left(\alpha_{SC} A^{n_{SC}} + \alpha_F E^{n_F} + \alpha_Z z^{n_z}\right). \tag{15}$$

The operational cost of the plant concerns thermal and electrical energy, consumed at heat exchangers and fans, respectively:

$$C_{OP} = \sum_{\text{Sections}} \left(C_E E + C_{ST} z\right). \tag{16}$$

On the basis of the preceding, the total annual cost of the plant can be expressed by means of the following equation:

$$C_T = e C_{CP} + t_{OP} C_{OP}. \tag{17}$$

Given a configuration for a conveyor-belt dryer, the design variables determining the total cost belong to a set of the form $\{T_{AC}, R, \Delta T_A, t\}$ for each section participating in the dryer. These variables can represent the process in a more straightforward way, due to their explicit physical meaning, and all other variables involved in the overall process model can be calculated accordingly.

In continuous fluidized bed dryers, the particulate solid phase is completely dispersed in a vertically flowing gas stream as a consequence of the buoyancy effect of the gas. Due to the agitation in the fluidized bed, good mixing of the solid phase is usually achieved. Moreover, the turbulent activity in the bed, produces high rates of heat transfer between the gas and solid phases, and results in a uniformity of the solid-phase temperature throughout the dryer. Fluidization exists when the gas superficial velocity varies between two extreme values, corresponding to fluidization and entrainment effects. The particular gas velocities are chiefly influenced by particle density and size. The drying air stream is composed of the combustion gases coming from a burner. It is common practice that temperature of the mixed solid–gas phase is controlled by means of the fuel flow rate. Furthermore, in some cases the outlet-air-stream humidity is controlled by the inlet-air-stream flow rate. A typical flow sheet comprising the previously mentioned structure, as well as the arrangement of its control facilities, is presented in Figure 1.

The steady-state mathematical model of a fluidized-bed dryer involves heat and mass balances as well as constraints based on construction reasoning. The burner is modeled exactly as in the conveyor-belt dryer case. The overall humidity and energy balances within the dryer are given by the following equations:

$$F_S(X_{S0} - X_S) = F_{AC}(X_A - X_{A0}) \tag{18}$$

$$F_S(h_S - h_{S0}) = F_{AC}(h_{AC} - h_A). \tag{19}$$

We again assume that the heat-transfer coefficient is high enough so that the product stream leaving the dryer is in thermal equilibrium with the air stream leaving the mixed solid–gas phase. Drying kinetics are similar to the ones derived for the conveyor-belt dryer. In this case, however, drying conditions affecting kinetics are the ones prevailing within the fluidized-bed bulk phase, assuming that complete mixing of phases takes place. Air flow rate within the dryer is given by the following equation as a function of the porosity of the fluidized bed:

$$F_{AC} = \rho_A A V_A \epsilon. \tag{20}$$

The corresponding bed height is given by the following relation, an alternative expression for solid holdup in the dryer:

$$H = \frac{F_S(1 + X_{S0})t}{\rho_S A(1 - \epsilon)}. \tag{21}$$

Fluidized-bed porosity is given by the Todes equation, and is recommended for related calculations on the entire fluidization region (Aerov and Todes, 1968; Strumillo and Kudra, 1986):

$$\epsilon = \left(\frac{18Re + 0.36Re^2}{Ar}\right)^{0.21}. \tag{22}$$

Pressure drop along the dryer is given as follows, where the gas flow through the fluidized bed is treated as an equivalent one through a porous body with given porosity (Marcus et al., 1990):

$$\Delta P = (\rho_S - \rho_A)(1 - \epsilon)\mathrm{Hg}. \tag{23}$$

This relation is suggested for the condition of incipient fluidization, but it can safely be used for the entire fluidization region since the pressure drop remains constant from fluidization up to entrainment (Hovmad, 1995). Fluidization is guaranteed if air velocity varies between two extremes. Incipient fluidization velocity is calculated by the following equation, which is derived from the theory of free settling:

$$V_A^{\min} = \left[\frac{4}{3}\frac{d_P(\rho_S - \rho_A)}{\rho_A}\frac{1}{(0.4 + 24/Re + 4/Re^{1/2})}\right]^{1/2}. \tag{24}$$

Strumillo and Kudra (1986) list over 20 equations for the determination of this value as suggested by various researchers. Romankov and Rashkovskaya (1979) recommend the following equation for the calculation of entrainment gas velocity:

$$V_A^{max} = V_A^{min}\left(0.1175 - \frac{0.1046}{1+0.00373Ar^{0.6}}\right). \tag{25}$$

On construction reasoning it is recommended (Strumillo and Kudra, 1986) that

$$H \leq 3D. \tag{26}$$

The economic evaluation of the dryer is based on the determination of its total annual cost. The corresponding capital cost is affected by its cross-sectional area, the installed power of the fan involved, and the flow rate of the fuel. All capital-cost components obey economy-of-scale laws:

$$C_{CP} = \alpha_{D,FB}A^{n_{D,FB}} + \alpha_F E^{n_F} + \alpha_Z Z^{n_z}. \tag{27}$$

The operational cost of the plant concerns thermal and electrical energy, consumed by fuel and fans, respectively:

$$C_{OP} = C_E E + C_Z Z. \tag{28}$$

Based on the preceding, the total annual cost of the plant can be expressed by Eq. 17. For a fluidized-bed dryer, design variables determining the total cost belong to a set of the form $\{A, V_A, T_{AC}\}$. All other variables involved in the mathematical model of the process can be calculated accordingly.

In continuous rotary dryers the free-flowing granular solids are passed through a rotating cylinder and are showered through a hot gas moving usually countercurrently to the solid phase. They involve a burner for direct heating of the product with hot combustion gases. It is common practice that the temperature of the inlet air stream is controlled by means of the steam or fuel flow rate. Furthermore, in some cases the outlet-air-stream humidity is controlled by means of the inlet-air-stream flow rate. Rotary dryers involve rather simple flow sheets. A typical flow sheet comprising the previously mentioned structure, as well as the arrangement of its control facilities, is presented in Figure 1.

The steady-state mathematical model of a rotary dryer involves heat and mass balances as well as constraints based on construction reasoning. The burner is modeled exactly as in the conveyor-belt-dryer case. The following differential equations involve mass and energy transfer within the product and air phase:

$$-F_S\frac{dX_S}{dt} = F_{AC}\frac{dX_A}{dt} \tag{29}$$

$$-F_S\frac{dX_S}{dt} = k_M X_S W \tag{30}$$

$$F_S\frac{dh_S}{dt} = U_A(T_A - T_S)W + F_S\Delta H_S\frac{dX_S}{dt} \tag{31}$$

$$F_{AC}\frac{dh_A}{dt} = U_A(T_A - T_S)W. \tag{32}$$

Drying kinetics are incorporated by means of Eq. 30. In Eqs. 31 and 32 a volumetric heat-transfer coefficient is introduced in order to model heat transfer appropriately (Douglas et al., 1993). Equations 29–32 have to be solved simultaneously. The product holdup of the dryer is given by the following equations as a function of the overall residence time:

$$W = F_S(1 + X_{S0})t \tag{33}$$

$$W = F_S(1 + X_{S0})t. \tag{34}$$

The volumetric coefficient, f_S, expresses the volumetric fraction of the solid phase in the total dryer volume. Its recommended value varies between 7 and 8% (Wentz and Thygeson, 1988). The following construction constraint (Wentz and Thygeson, 1988) is suggested:

$$D \leq 5L. \tag{35}$$

The temperature diminution of the drying air stream in passing through the solid particles is given as follows:

$$\Delta T_A = T_A - T_{S0}. \tag{36}$$

Its value should be greater than a minimum value that would guarantee satisfactory heat transfer. The economic evaluation of the dryer is based on the determination of its total annual cost. The corresponding capital cost is affected by the peripheral surface area of the dryer, the installed power of fan involved, and the fuel flow rate. All capital-cost components obey economy-of-state laws:

$$C_{CP} = \alpha_{D,RD}A_P^{n_{D,RD}} + \alpha_F E^{n_F} + \alpha_Z Z^{n_Z}. \tag{37}$$

The operational cost is given by Eq. 28. Again, the total annual cost of the plant can be expressed by Eq. 17. For a rotary dryer, the design variables determining the total cost belong to a set of the form $\{A_P, V_A, T_{AC}\}$. All other variables involved in the mathematical model of the process can be calculated accordingly.

Specific enthalpies of product and air streams can be calculated by the following equations (Pakowski et al., 1991):

$$h_S = c_{PS}T_S + X_S c_{PW} T_S \tag{38}$$

$$h_A = c_{PA}T_A + X_A(\Delta H_0 + c_{PV}T_A). \tag{39}$$

These properties are dealt with as linear functions of temperature and water content, since the corresponding specific heats of solid particles, dry air, water, and vapor are assumed to be constant within the desired temperature range.

Design of Convective Industrial Dryers

On the basis of the preceding, the design strategy for a dryer type studied can now be clearly stated. Given the specified product characteristics with a predefined flow rate, to be dried from an initial to a desired moisture content level, under constraints imposed by construction and product quality reasoning, the following must be determined:

1. The optimum configuration when this is not clear-cut, that is, for conveyor-belt dryers the number of drying sections (flow-sheet structure).
2. The appropriate sizing of the equipment (construction characteristics).
3. The best setpoints of controllers (operating conditions).

For the operational analysis of the process, the structure and sizing characteristics of the equipment are predefined. In this case, we seek the economic performance of a specified dryer when it is operated under different operating conditions, that is, various production capacities, flexibility with respect to input variables, and so forth. In this way, design vs. operational performance is studied and various dryer types can be compared in detail for selection purposes.

In addition to the analysis on the specific dryer types, the product characteristics must be taken into consideration. In this way, more general aspects of the problem can be obtained. The product is generally characterized by its nominal moisture content, its desired moisture content level after the completion of drying, that is to say the water loss defined as the ratio of the target to the initial moisture content, and its drying characteristics. The most important characteristic affecting drying kinetics of the product is its corresponding effective diffusion coefficient, which strongly depends on temperature through the following correlation (Marinos-Kouris and Maroulis, 1995):

$$D_e = D_0 \exp(-E_A/RT). \quad (40)$$

An extensive review of the literature in this field revealed that the dependence of effective moisture diffusivity on temperature involves an average activation energy of the diffusion value of 35 kJ/mol for most systems studied within a range of 10% (Marinos-Kouris and Maroulis, 1995). The most important fact is that its corresponding Arrhenius factor, D_0, varied within values ranging from 10^{-3} to 10^{-5} m²/s. In this way we can describe the kinetics of most physical systems undergoing drying using only two parameters, from which one is approximately constant for all cases studied. In the previous analysis of dryer models, however, we utilized the drying constant concept. Unlike moisture diffusivity, which has physical meaning and describes moisture diffusion within a porous medium, the drying constant is a phenomenological term incorporating all transfer mechanisms and has empirical origin. In theory, it is impossible to estimate an empirical constant using theoretical reasoning. Nevertheless, if we assume that the controlling mechanism of the phenomenon is moisture diffusion, then the drying constant can be expressed as a function of moisture diffusivity through the expression (Marinos-Kouris and Maroulis)

$$k_M = \pi^2 D_e/d_P^2. \quad (41)$$

This relation reveals the dependence of the drying constant on particle-characteristic dimensions. These remarks conclude our effort for a general as well as accurate description of the product characteristics. Apart from the production capacity of each dryer, the drying-air-stream temperature level must also be taken into consideration. Various quality reasons (thermal degradation of ingredients) dictate that a wide variety of products (chiefly food materials) be dried under low to moderate temperature levels (less than 80°C). Products of a basically inorganic nature are dried to temperature levels exceeding 400°C.

By using the analysis described so far, we could design dryer structures and compare rival types by obtaining the corresponding optimal cost values through optimization manipulations of the process model for each dryer type examined, provided that the following are specified:

1. Production capacity of the plant (10 – 10,000 kg/h dry basis (db)).
2. Drying temperature level (80 – 400°C).
3. Initial material moisture content (0.5 – 5 kg/kg db).
4. Water loss (50 – 99%).
5. Arrhenius diffusivity factor (10^{-3} – 10^{-5} m²/s).

In this analysis of the problem, both product and dryer characteristics are treated simultaneously so that generalized design curves can be obtained. We have assumed raw-material particles with a characteristic dimension of 10 mm and a bulk density of 1,500 kg/m³. Fresh air is available at 0.01 kg/kg db humidity and 25°C temperature conditions. The product enters the drying system at fresh-air temperature. Uniform drying in conveyor-belt dryers is achieved by not allowing drying-air-temperature diminution through the product to exceed a value of 10°C. The same temperature level is assumed as the minimum approach for the rotary dryers where a volumetric heat-transfer-coefficient value of 30 kW/kg·K db was used. The maximum belt load allowed by construction specifications for the case of conveyor-belt dryers is 50 kg/m² wet basis (wb). The preferable recycle ratio for the dryer is 10. The cost of a conveyor-belt is \$1,600/m² increased by a power law 0.75 as area varies. Each section contributes an additional cost of \$1,400/m², with a corresponding 0.3 power law per section area considered. The figures for fluidized bed and rotary dryers are \$1,600/m² and \$8,000/m², varied by corresponding power laws of 0.75 for both cases, for the cross and peripheral surface area, respectively. The cost of the burner utilized is \$200/kg over one-hour basis operation, increased by a power law of 0.4 as fuel flow rate varies. Fans add a capital cost of \$500/kWh, increased by a 0.3 power law. Electrical energy costs 8¢/kWh. The cost of fuel is 20¢/kg, its hydrogen concentration is 0.15 kg/kg, and its heat of combustion is 40,000 kJ/kg. The drying system is operated on the basis of 2,000 h/yr, and the capital cost will be paid off within a period of 5 years.

The optimization procedure for the determination of the optimal cost structures for all cases and processes and economic figures described earlier was concentrated on the solution of the NLP problems described in the modeling section. Nonlinear constrained optimization throughout this article was carried out by means of the successive quadratic programming algorithm in the form of subroutine E04UCF/NAG. The differential equations of the rotary dryer were solved by approximately discretizing the control volume and solving the resulting algebraic equations by a Newton-based iterative routine (Kamke and Wilson, 1986). The code was executed on an SG Indy workstation under Unix.

The results of the optimization procedure for the conveyor-belt dryers are given in the diagrams of Figure 2, where the optimual cost is plotted against production flow rate for various values of process and product characteristics. We note that a multiple-section dryer is preferred in the cases where drying conditions are more difficult to handle, that is, low temperature levels and Arrhenius diffusion factors as well as high water loss specifications. Multiple sections tend to be used even at lower initial moisture content values for the same drying conditions as the ones just mentioned. At higher temperature levels, however, production capacity tends to be the only substantial process variable affecting cost. In this case all curves are grouped together in families located very close to one another. The same procedure was carried out for the remaining dryer types, that is, fluidized-bed and rotary dryers, and are given in the diagrams of Figures 3 and 4, respectively. The generalized curves of these diagrams are very important to the design. They contain significant concentrated information for all types of dryers and product characteristics examined. In particular, they are quite important when all dryer types are to be explicitly compared. At low temperature levels, fluidized-bed dryers appear to be cheaper than rotary dryers, with costs comparable to the ones of the conveyor-belt-type designs. At higher temperature levels, however, these two types exhibit much cheaper designs with respect to the conveyor-belt results, indicating that this type of dryer is not suitable for the drying of inorganics. In all cases, however, fluidized-bed dryers tend to be more flexible, revealing a wide range of design characteristics for each individual specification.

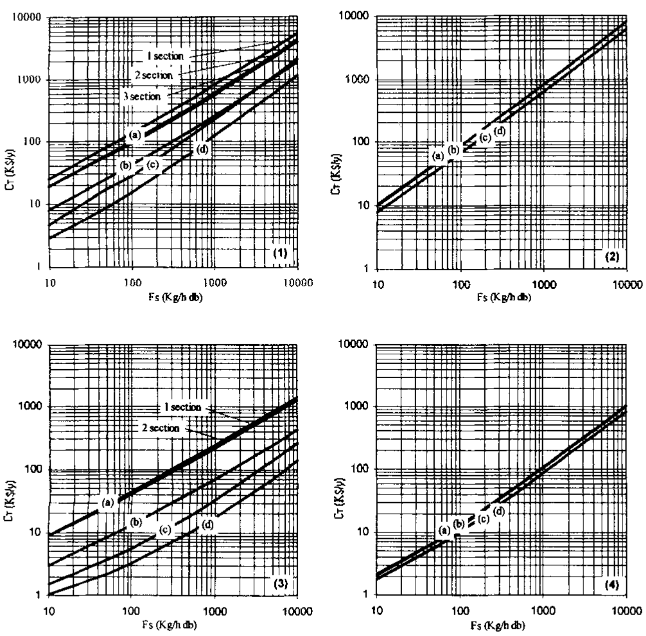

Figure 2. Total optimal cost against production capacity for conveyor-belt dryers.

(1): $X_{S0} = 5$ kg/kg db, $T_{AC} = 80°C$; (2): $X_{S0} = 5$ kg/kg db, $T_{AC} = 400°C$; (3): $X_{S0} = 0.5$ kg/kg db, $T_{AC} = 80°C$; (4): $X_{S0} = 0.5$ kg/kg db, $T_{AC} = 400°C$; (a): $D_0 = 10^{-5}$ m²/s, $X_S/X_{S0} = 99\%$; (b): $D_0 = 10^{-5}$ m²/s, $X_S/X_{S0} = 50\%$, (c): $D_0 = 10^{-3}$ m²/s, $X_S/X_{S0} = 99\%$ (d): $D_0 = 10^{-3}$ m²/s, $X_S/X_{S0} = 50\%$.

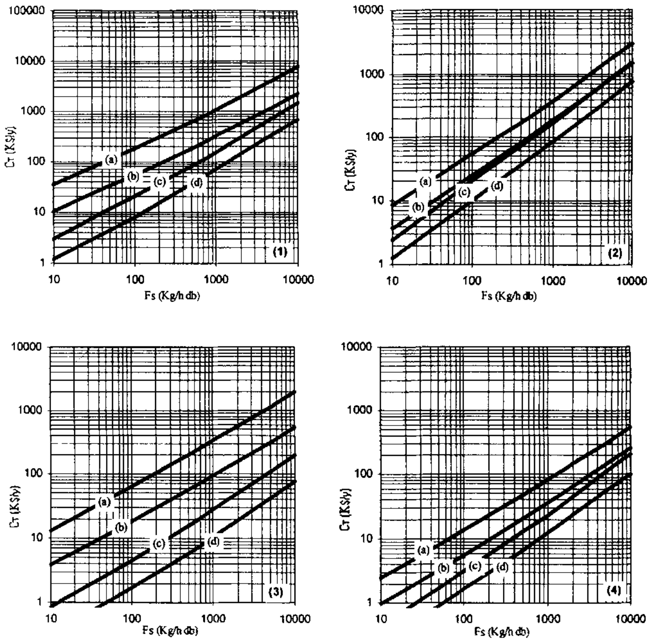

Figure 3. Total optimal cost against production capacity for fluidized-bed dryers.

Symbols as in Figure 2.

Case Studies

The proposed methodology was applied to the design of a dehydration plant that treats 1,800 ton/yr wb of vegetables (potatoes), that is, 150 kg/h db. A raw material, with an initial material moisture content level of 5 kg/kg db, is to be dried at a desired material content level of 0.05 kg/kg db (water loss 99%). Its Arrhenius diffusivity factor is 0.0004

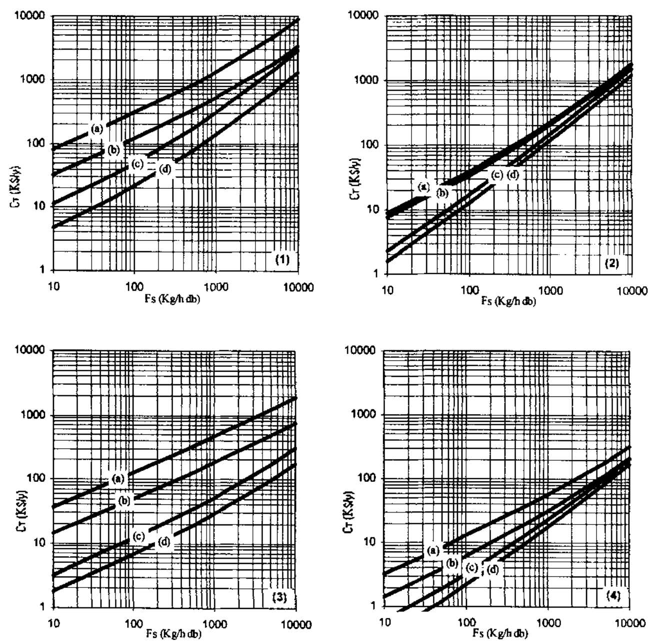

Figure 4. Total optimal cost against production capacity for rotary dryers.

Symbols as in Figure 2.

m^2/s. For thermal degradation reasons, the product is not to be dried at temperatures exceeding 80°C. The best conveyor-belt dryer structure evaluated utilized a two-section dryer that involved 34 m^2 and 17 m^2 conveyor-belt areas for each section. The drying-air-stream flow rates were 144,000 m^3/h and 25,000 m^3/h db, respectively. Fuel consumption for the two sections was 62.2 kg/h and 10.8 kg/h. The total cost was \$38,400/yr and involved a 68% operational component (2.1¢/kg wb for the product treated). The corresponding design for the fluidized-bed dryer involves a 2.2-m-dia. structure. The drying-stream flow rate is 55,700 m^3/h db, the outlet product and air temperature are 57.8°C, and fuel consumption is 179.7 kg/h. The total annual cost is \$44,200/yr, with an operational component of 87% (2.4¢/kg wb for the product treated). The rotary dryer design resulted in a 13.7-m-length dryer involving 40,000 m^3/h db of drying-stream flow rate and a 61.5 kg/h fuel consumption. The total annual cost is \$57,200/yr, with an operational component of 28% (32¢/kg wb for the product treated).

The effect of the variations of selected dryer parameters on the predicted optimal cost for the case of vegetables and for all the dryer types studied is given in the diagrams of Figure 5. The base conditions for all comparisons are the ones evaluated by the optimization procedure. Each variation from the base condition, plus or minus 20%, was computed while all other conditions remained constant. Within the range of conditions examined, product flow rate and initial material moisture content had the greatest effect on the predicted optimal dryer cost. In the particular case of the rotary dryer, the effect of drying-stream temperature and particle characteristic size was significant as well.

Suppose a dryer is constructed based on specific demands for its production capacity. It is common experience that the plant is designed so that its dryers can operate under various production planning platforms based on market demands. Therefore, it is expected that particular purchased equipment could meet various production constraints and work under different product flow rates. In the case where the flowsheet structure and its corresponding construction characteristics are specified, optimal operational parameter levels can be found by minimizing the dryer operational cost. If one elects a nominal product flow rate for dryer operation, then the resulting optimal operation problem can be dealt with by means of the mathematical programming approach mentioned in the previous sections, for all dryer types involved. Since all construction variables are given, the number of design variables that remain are substantially reduced and, therefore, the resulting NLP problem is significantly simplified. In this case, we are able to express the optimal operational cost as a function of product flow rate, for a given flow-sheet configuration. If the production capacity is considered to vary, then the corresponding optimal operational cost can be estimated for a given structure and, therefore, a great number of related flow-rate–optimal-operational-cost pairs can be obtained. The corresponding flow-rate–optimal-cost curves for the vegetables case is given in Figure 6. These curves are also considered to be extremely important, since they include concentrated information for the dryer's operational behavior. The shape of the curves exhibit a linear performance for small values of product flow rate, up to a certain production level. From then on, the operational cost is greatly increased until the maximum production level is reached. Clearly, the proposed structures are unable to process the specified product when the production level exceeds this extreme value. The fluidized-bed dryer's optimum operational cost is kept high even at a negligible production flow-rate level. This is due to the fact that the drying-air flow rate must remain high in order to achieve fluidization, even for extremely low product holdup. In the corresponding design curves, this effect was not observed; dryer diameter was left to vary, being evaluated by the optimization procedure. In the curve of Figure 6, however, the dryer diameter is con-

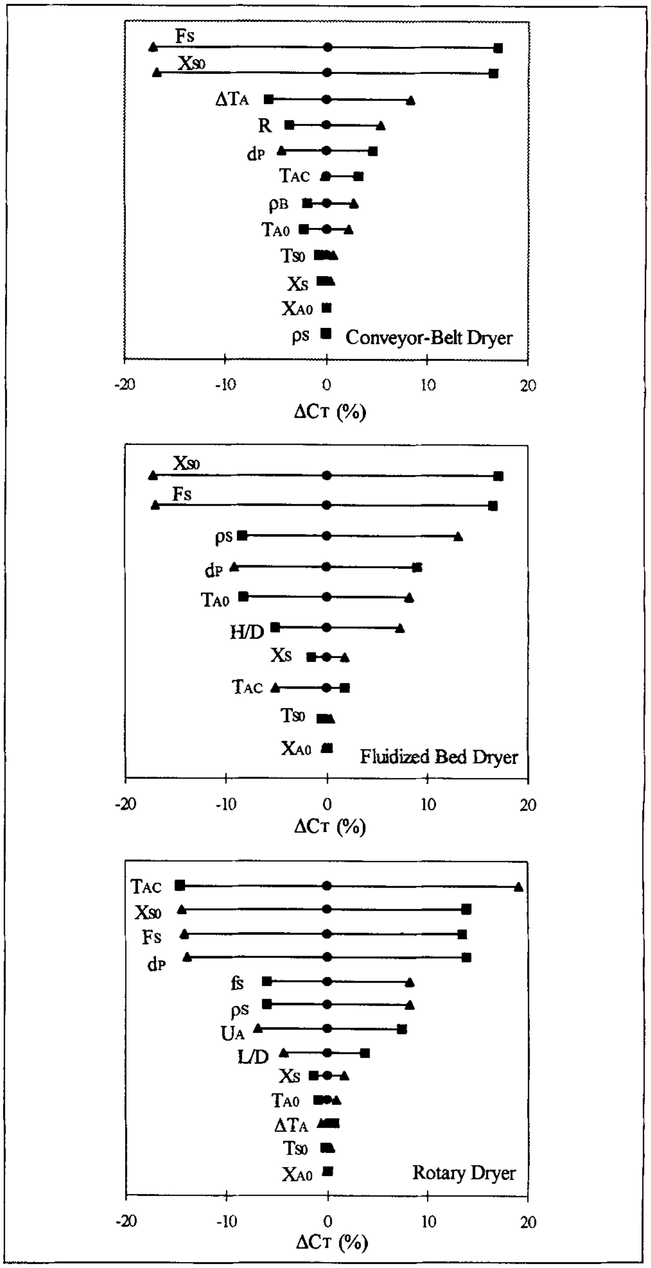

Figure 5. Effect of variations of selected convective dryer parameters on the predicted total optimal cost for the 150 kg/h db case of foods for all dryers examined.

▲: −20% of parameter, ■: +20% of parameter, ●: base design case.

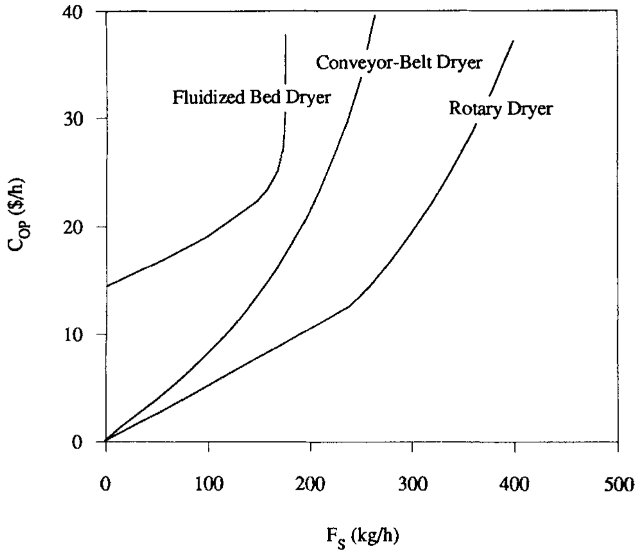

Figure 6. Optimal operational cost against production capacity for all convective dryers of the foods case.

stant, resulting in high drying-air flow rates due to fluidization constraints. The rotary dryer can be used for very high levels of production capacity (more than 250% of the nominal value), while the fluidized one reaches its maximum production level rapidly for a value of only 14% of the nominal one.

Cost curves are the essence of the dryer selection procedure adopted. Design is compared to operational performance to reach a final decision. Rotary dryers involve expensive capital equipment, but guarantee quite satisfactory operation. Fluidized-bed dryers involve huge operational expenses and are not flexible in operation; however, they are very cheap dryers. Conveyor-belt dryers in the case of food products are probably the best solution. The results of the analysis adopted are based on strictly economical and technical criteria. Obviously, different cost scenarios could produce completely different results.

As a second case study we choose to design a dehydration plant that treats 30,000 ton/yr wb of inorganic materials (bentonite), that is, 10,000 kg/h db. Bentonite is a significant mineral product in Greece, whose main uses are as pelletizing material in drilling muds, in foundries, and as cat-litter, and is mined chiefly on Milos. The raw material, with an initial material moisture content level of 0.5 kg/kg db, is to be dried at a desired material content level of 0.25 kg/kg db (water loss 50%). Its Arrhenius diffusivity factor is 0.001 m^2/s. The product is to be dried at very high temperature levels (400°C). The design for the fluidized-bed dryer involves a 1.2-m-dia. structure. The drying-stream flow rate is 17,000 m^3/h db, the outlet product and air temperature is 180°C, and the fuel consumption is 179 kg/h. The total annual cost is $76,900, with an operational component of 88% (0.25¢/ton wb for the product treated). The rotary dryer design resulted in a 2.7-m-long dryer involving 37,000 m^3/h db of drying-stream flow rate and a 404.2 kg/h fuel consumption. The total annual cost is $138,000 with an operational component of 45% (0.46¢/ton wb for the product treated). The same design for a conveyor-belt dryer involved a total cost of $832,000/yr; obviously using such a dryer is completely out of the question.

The effect of variations of selected dryer parameters on the predicted optimal cost for inorganics and for both dryer types studied is given in the diagrams of Figure 7. The conditions studied were similar to the ones for Figure 5. Within the range of conditions examined, the product flow rate, the initial, and desired material moisture content had the greatest effect on the predicted optimal dryer cost. In this case the effect of these variables on the estimated optimal cost was even greater.

In the case of bentonite, the operational problem discussed in the food products case is more intense. Different final uses of the product require different specifications related to product moisture content. For example, feed stock concentration in montmorillonite and material moisture content. Thus, design is of less significance than the effort required to actually operate the dryers under different production specifications. Therefore, it is expected that specific purchased equipment could meet various production constraints and work under different product flow rates. In the case

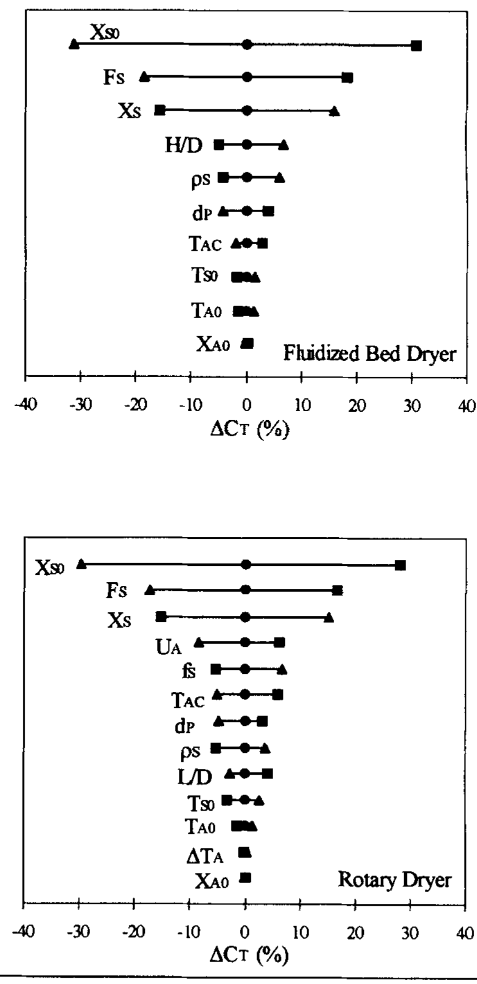

Figure 7. Sensitivity analysis of the optimum 10,000 kg/h db case of inorganics for all dryers examined.

Symobls as in Figure 5.

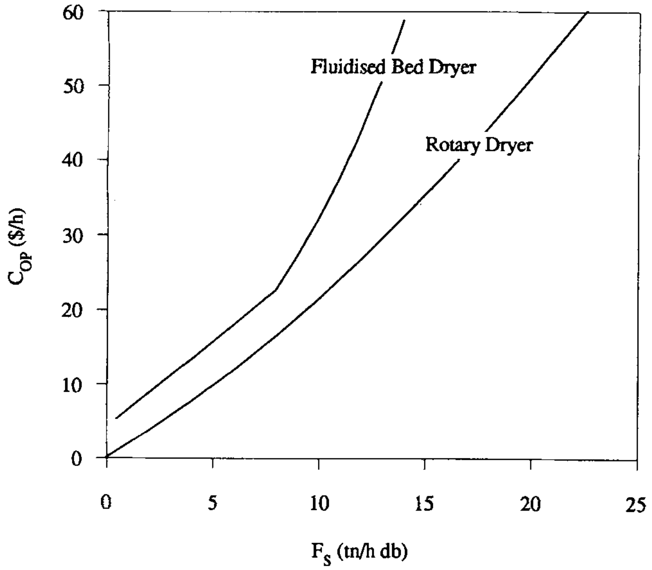

Figure 8. Optimal operational cost against production capacity for all convective dryers of the inorganics case.

where the flow-sheet structure and its corresponding construction characteristics are specified, optimal operational parameter levels again can be evaluated by minimizing the dryer operational cost. If the production capacity is considered to vary, then the corresponding optimal operational cost can be estimated for a given structure and, therefore, a great number of related flow-rate–optimal-operational-cost pairs can be obtained. For the nominal case for all dryers evaluated in the previous sections, the corresponding flow-rate–optimal-cost curves are given in Figure 8. The shape of the curves exhibit a linear performance for small values of product flow rate, up to a certain production level. From then on, the operational cost is greatly increased until the maximum production level is reached. The fluidized-bed dryer optimum operational cost is kept high even at negligible production flow-rate levels, as explained in the food products case. The rotary dryer can be used for very high levels of production capacity (more than 120% of the nominal value), while the fluidized-bed one reaches maximum production level for a value 50% of the nominal one, that is, the overall performance is better than in the food products case. In Figures 2–4 and 8, design is compared to operational performance in order to reach a final decision. Rotary dryers involve expensive capital equipment, but guarantee quite satisfactory operation, while fluidized-bed dryers involve huge operational expenses, but are very cheap to construct. Again, the results of the analysis adopted are based on strictly economical and technical criteria, on the basis that different cost scenarios can produce completely different results.

Conclusion

The dryer selection procedure can be substanitally aided by evaluating the design vs. operational performance characteristics for various different dryer types. The design aspects can be successfully explored by constructing the optimal total annual cost curves for a wide range of product and process characteristics. These curves contain concentrated information on design with respect to optimum flow-sheet structure, sizing, and operational parameters. The process could be adequately described by deducing its *mathematical model*. When a design configuration is implemented, its operational performance can be investigated by suitably constructing optimal operational cost curves over a wide product flow-rate range. These curves reveal the *a priori* performance of various product demand scenarios imposed by the market. This methodology was conveyed to the continuous convective dryer's case. Three different types of dryers were investigated with regard to design; namely, conveyor belt, fluidized bed, and rotary. Rotary dryers turned out to be rather expensive compared to fluidized-bed dryers, but it is the other way around when operational performance is taken into consideration. Conveyor-belt dryers, when utilized, lie somewhere in between, producing satisfactory results with respect to both design and operation. This methodolgy can be safely applied to a wide range of materials, from food products to inorganic minerals.

Notation

A = area of cross section of dryer, m^2
A_P = area of peripheral surface of dryer, m^2
Ar = Archimedes number
C_E = cost of electricity unit, $/kWh
C_H = concentration of fuel in hydrogen, kg/kg
C_{PA} = specific heat of air, kJ/kg·K
C_{PS} = specific heat of dry solid, kJ/kg·K
C_{PV} = specific heat of vapor, kJ/kg·K
C_{PW} = specific heat of water, kJ/kg·K
C_Z = cost of fuel, $/kg
D = diameter of the dryer, m
d_P = particle diameter, m
e = percentage of capital cost on an annual rate
E_A = activation energy for diffusion, kJ/mol
F_A = flow rate of fresh air stream, kg/h db
F_{A0} = flow rate of input air stream, kg/h db
F_S = flow rate of product stream, kg/h db
g = gravitation constant, m/s^2
L = length of dryer, m
$n_{D,CV}, n_{SC}, n_F, n_Z$ = capital cost constants
P = total pressure, kPa
Re = Reynolds number
t_{OP} = operating time for dryer operation, h/yr
T_A = temperature of output-air stream, °C
T_{AC} = temperature of drying-air stream, °C
T_{AM} = temperature of mixed-recirculation and fresh-air streams, °C
T_{A0} = temperature of fresh-air stream, °C
T_{S-} = temperature of product stream on entering the section, °C
T_{S+} = temperature of product stream on leaving the section, °C
U_A = volumetric heat-transfer coefficient, kW/kg K
V_A = air velocity through product, m/s
X_A = absolute humidity of output air stream, kg/kg db
X_{AC} = absolute humidity of drying air stream, kg/kg db
X_{A0} = absolute humidity of fresh-air stream, kg/kg db
X_{S0} = nominal value of material moisture content entering the dryer, kg/kg db
X_{S-} = moisture content of product stream on entering a drying section, kg/kg db
X_{S+} = moisture content of product stream on leaving a drying section, kg/kg db
z = flow rate of fuel, kg/h

Greek letters

$\alpha_{D,CV}, \alpha_{SC}, \alpha_F, \alpha_Z$ = capital cost constants
ΔH_S = latent heat of vaporization of water, kJ/kg

ΔH_F = heat of combustion of fuel, kJ/kg
ΔH_0 = latent heat of vaporization of water at reference temperature, kJ/kg
ρ_A = density of air, kg/m^3
ρ_S = bulk particle density, kg/m^3 wb

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Manuscript received Nov. 28, 1995, and revision received Apr. 8, 1996.