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AN INVESTIGATION INTO THE RELATIONSHIP OF FLUIDIZATION VELOCITY TO THE OPTIMAL ECONOMIC DESIGN OF A FLUIDIZED-BED DRYING PROCESS

THESIS

DEPARTMENT OF THE AIR FORCE
AIR UNIVERSITY (ATC)
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Wright-Patterson Air Force Base, Ohio

AFIT/GOR/OS/82D-10

AN INVESTIGATION INTO THE

RELATIONSHIP OF FLUIDIZATION

VELOCITY TO THE OPTIMAL ECONOMIC

DESIGN OF A FLUIDIZED-BED

DRYING PROCESS

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Moussa I.M. Mostafa
Lt. Col. Egypt Army

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THESIS

Presented to the Faculty of the School of Engineering
of the Air Force Institute of Technology
Air University
in Partial Fulfillment of the
Requirements for the Degree of
Master of Operations Research

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by
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Graduate Operations Research

December 1982

Preface

The use of the fluidized solids technique was developed very largely by the petroleum and chemical industries, for processes where the very high heat transfer coefficients and the high degree of uniformity of temperature within the bed enabled the development of processes which would otherwise be impracticable.

Fluidized solids are now used quite extensively in many industries where it is desirable to bring about intimate contact between solid particles and gas stream. The principles of fluidization are applied in the drying processes to get a fluidized-bed drying operation.

The purpose of this study is to develop an analytical model of the fluidized-bed drying process which show the behavior of the capital and power cost of such a system in response to changes in fluidization velocity.

I would like to express my gratitude to my thesis advisor Professor Robert F. Allen for his most valuable advice and guidance during this study. I would also like to thank Professor Milton Franke, my reader, for his help during the study.

The demanding nature of any thesis project is hard on the student accomplishing the thesis, but it is equally as hard on the spouse of the student. I am indebted to my wife, Dr. Bousiana, for her constant support throughout this project.

MOUSSA I. M. MOSTAFA

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ABSTRACT

This thesis investigates the economic characteristics of a fluidized-bed drying process. It focuses on the way in which fluid velocity impacts the capital and power consumption costs of such a system.

The engineering and economic relationships employed in the model are developed and a sensitivity of the optimal design of a fluidized-bed drying process based on capital and operating cost is explored through variation in the technical parameters of the model.

NOMENCLATURE

English Letter Symbols

A_c	Cross-sectional area of the dryer, sq. ft.
A_H	Peripheral area of the heater, sq. ft.
A_P	Peripheral area of the dryer, sq. ft.
C_s	Humid heat capacity of one lb of dry air and the moisture it contains
dA_H	Element of surface area, sq. ft.
D	Dryer diameter, ft.
D_P	Particle diameter, ft.
F	Gas flow rate, lb. dry air/hr.
g	Acceleration of gravity, 32.2 ft./(sec.) (sec.)
g_c	Conversion factor, 32.2 (lb.) (ft.)/(lb_f.) (sec.) (sec.)
G	Mass flow rate of gas in the dryer, lb./(sq. ft.) (hr.)
G_{mf}	Fluid mass velocity at minimum fluidization, lb./(sq. ft.) (Sec.)
$G(u)$	Objective function
H	Absolute humidity, lb. water/lb. dry air
H_{ad}	Adiabatic head, ft.
H_i	Moisture content of entering air, lb. water/lb. dry air.
H_o	Moisture content of leaving air, lb. water/lb. dry air.
i	Average rate of return on invested capital, (frac./yr.)
i_m	Minimum acceptable return rate on invested capital.
K	Proportionality constant.
k	Specific heat ratio for air, dimensionless.

L	Length of the dryer, ft.
L_r	Supscript for large size equipment.
M	Maintenance charge, (frac./yr.)
n	Constant.
n_1	Equipment size-cost exponent.
n_2	Allowable life for depreciation of equipment.
P	Pressure, lb_f./sq. in.
P_1	Entering gas pressure, lb_f/sq. in.
P_2	Compressed gas pressure, lb_f/sq. in.
Q	Flow rate, cu. ft./min.
Qa	Rate of heat transfer in the dryer, B.t.u./hr.
Q_H	Total heat transfared, B.t.u./hr.
r	Fixed charge factor, (frac./yr.)
R	Gas constant, ft./(lb. mole) (^{o}K)
R_e	Reynolds number, dimensionless.
S	Supscript for small size equipment.
t	Income tax rate, (frac./yr.)
T_{Di}	Inlet air dry bulb temperture, ^{o}F.
T_{Do}	Outlet air dry bulb temperature, ^{o}F.
T_{gi}	Inlet air temperature, ^{o}F.
T_{go}	Outlet air temperature, ^{o}F.
T_s	Steam temperature, ^{o}F.
T_w	Air wet bulb temperature, ^{o}F.
u_{mf}	Minimum fluidization velocity, ft./sec.

u_t	Terminal velocity, ft./sec.
U_a	Overall volumetric heat transfer coefficient, B.t.u./(hr.) (cu. ft.) (^oF).
U_H	Overall heat transfer coefficient, B.t.u./(hr.) (sq. ft.) (^oF).
w	Gas weight rate, lb./sec.
W	Pounds of dry solid, lb./hr.
v	Gas volume.
V	Dryer volume, cu. ft.
x_i	Entering solid moisture content, lb. water/lb. dry air.
x_o	Leaving solid moisture content, lb. water/lb. dry air.

Greek Letter Symbols

ΔP	Pressure drop, lb_f./sq. in.
ΔT_H	Logarithmic mean temperature difference between air and steam, ^oF.
ΔT_m	Logarithmic mean temperature difference, ^oF.
μ	Fluid viscosity, lb./(ft.) (sec.)
π	Constant
ρ	Air density, lb./cu. ft.
ρ_g	Gas density, lb./cu. ft.
ρ_s	Particle density, lb./cu. ft.
ϕ_s	Particle shape factor, dimensionless.

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AN INVESTIGATION INTO THE RELATIONSHIP OF FLUIDIZATION VELOCITY TO THE OPTIMAL ECONOMIC DESIGN OF A FLUIDIZED-BED DRYING PROCESS

I. INTRODUCTION

Fluidization may be defined as the phenomena in which the gravitational force acting on a dense swarm of particles is counteracted by the drag of an upward fluid stream, which cause these particles to be kept in a more or less floating state (14).

A. The Phenomenon of Fluidization

The term fluidization is used to describe a certain mode of contacting granular of finely divided solid particles with fluids. The physical phenomena occuring in the passage of fluids through unrestrained particles beds is described by many investigators. Zenz and Othmer (18), Leva (11), Kunii and Levenspiel (8), Coulson and Richardson (3), Lewis and Bowerman (13), Ergun (5) and Brownell and Katz (2).

Passing a fluid upward through a bed of fine particles involves certain characteristic states. These are shown in Figure 1 . When the flow rate is low, fluid percolates through the void space between stationery particles. This is the state of a "fixed bed".

With an increase in flow rate, an "expanded bed" occurs. At some velocity the particles are all just suspended in the upward floating gas or liquid. At this point the frictional

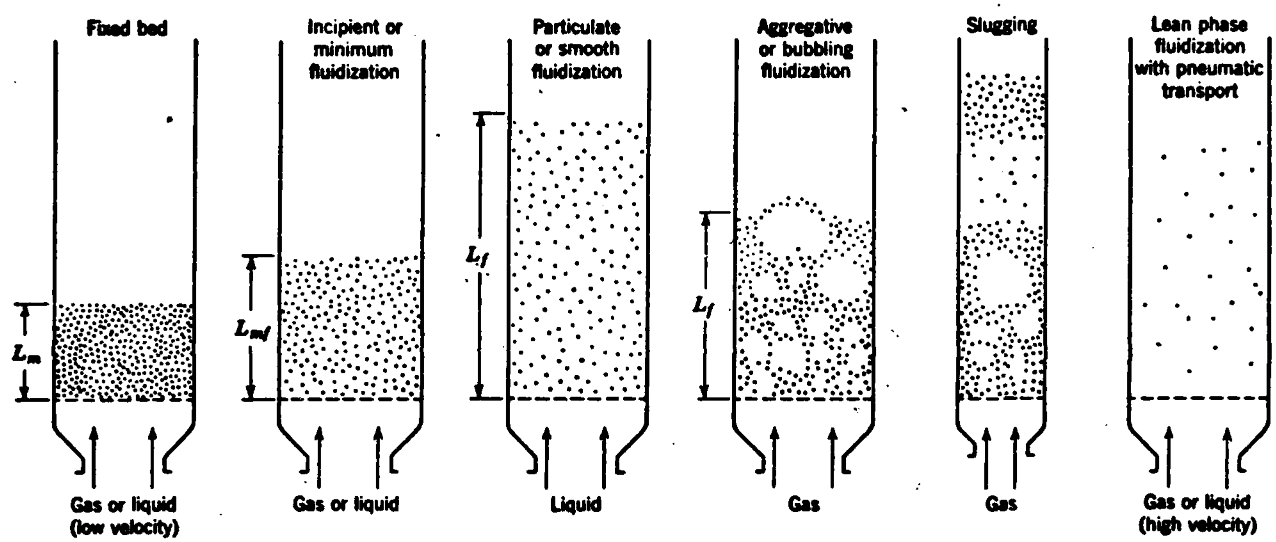

Figure 1 Various kinds of contacting of a bath of solids by fluid.
Kunii and Levenspiel (8)

force between a particle and fluid counterbalances the weight of the particle. The pressure drop through any section of the bed approximately equals the weight of fluid and particles in that section. The bed is referred to as an "incipiently fluidized bed" or a "bed at minimum fluidization" (8).

With an increase in flow rate beyond minimum fluidization, gas-solid systems behave in an unstable manner with bubbling and channeling. At higher flow rates, the movement of solids becomes more vigorous. Such a bed is called an "aggregative fluidized bed".

At a sufficiently high fluid flow rate the terminal velocity of the solids is exceeded, the upper surface of the bed disappears, entrainment becomes appreciable, and solids are carried out of the bed with the fluid stream.

The fluid bed reactor has a broad usage in chemical engineering industry (8). However, the complex nature of this fluid bed makes accurate mathematical modeling difficult. For example, Dale (5) describes the problem of waste disposal in the Navy in the late sixties and the Navy's interest in the combustion of solid material in a fluidized-bed reactor. The lack of fundamental design knowledge at that time was prohibiting accurate design and control of such a system. Dale (4) had limited success in building a mathematical model for non-catalytic reactions in a fluidized-bed reactor.

This study follows the spirit of the Dale research in that it represents a further exploration of an analytical model for fluidized-bed operations. However, the goal of this

research study is to investigate the economic characteristics of a fluidized-bed drying process. In particular, this study will focus on the way in which fluid velocity impacts the capital and power consumption costs of such a system. If successful this model can be used as a design tool in studies of fluidized-bed drying systems.

B. Objectives

The specific objective of this work is to develop an analytical model of the fluidized-bed drying process, which will highlight the behavior of the capital and power cost of such a system in response to changes in fluidization velocity. We are especially interested in determining the extent to which minimum system cost is sensitive to variation in fluidization velocity. A secondary objective of this study is to set out the cost-velocity relationships of each of the major equipment components of the fluidized-bed drying system.

C. Methodology

The method of approach used for this investigation is analytical. Using the available literature on chemical engineering process economics and cost estimation, a model is developed relating fluid velocity of the fluidized-bed drying operation to the capital and power consumption cost of the system.

This is done in two steps. First the basic engineering relationships pertaining to heat transfer, pressure change, equipment volume and surface area of equipment are specified

so as to highlight the role of fluid velocity. Second, these engineering parameters are related to capital and power consumption costs based on the empirical estimates of equipment cost size relationships. The resulting aggregate cost function represent an objective function to be minimized subject to a velocity range constraint. The upper and lower bounds of this velocity constraint are the terminal and minimum fluidization velocities, respectively.

The optimization technique used to solve the cost model is the Fibonacci search method (9).

II. ENGINEERING AND ECONOMIC ASPECTS FOR FLUIDIZED-BED DRYING

This chapter will summarize the key theoretical engineering relationships and economic cost concepts which will be used in developing of the aggregate cost model of fluidized-bed drying operations. The engineering relationships are mainly analytical equations which describe the physical operation, while the primary economic cost concepts are the equipment cost-size relationships and its conversion into annual costs by use of an annual capital charge factor.

A. Engineering Aspects

Fluidized Beds

At minimum fluidization, values of mass velocity and voidage can be related by Leva (3).

$$G_{mf} = \frac{0.005\ D_p^2\ g\ \rho_g(\rho_s - \rho_g)\ \phi_s^2\ \varepsilon_{mf}^3}{\mu(1 - \varepsilon_{mf})} \qquad (1)$$

Where G_{mf} = fluid superficial mass velocity for minimum fluidization, lb./(sec.) (sq. ft.);

D_p = particle diameter, ft.;

g = local acceleration due to gravity, 32.2 ft./(sec.)(sec.)

ρ_g = gas density, lb/cu. ft.;

ρ_s = solids density, lb/cu. ft.;

ϕ_s = particle shape factor, dimensionless;

ε_{mf} = voidage at minimum fluidization, dimensionless;

μ = fluid viscosity, lb./(ft.) (sec.)

If ε_{mf} and/or ϕ_s are unknown, a modification made by Wen and Yu (15) can be used which simplifies the expression of minimum fluidization velocity. Wen and Yu found that for a wide variety of systems:

$$\frac{1}{\phi_s \varepsilon_{mf}^3} \simeq 14 \text{ and } \frac{1 - \varepsilon_{mf}}{\phi_3^2 \ \varepsilon_{mf}^2} \simeq 11 \qquad (2)$$

which when substituted in the minimum fluidization velocity expression given by Kunii and Levenspiel (8) gives:

$$u_{mf} = \frac{D_p^2 \ (\rho_s - \rho_g) \ g}{1650 \ \mu} \qquad (3)$$

where u_{mf} = minimum fluidization velocity, ft/sec.

The gas flow rate through a fluidized bed is limited on the one hand by Umf and on the other hand by entrainment of solids by the gas. When entrainment occurs these solids must be recycled or replaced by fresh material to maintain stead-state operations. The upper limit of the gas velocity is approximated by the terminal or free-fall velocity of the particles which is reported by Kunii and Levenspiel (8) to be given by:

$$u_t = \frac{g \ (\rho_s - \rho_g) \ D_p^2}{18\mu} \text{ for Re} < 0.4$$

and

$$u_t = \left[\frac{4 \ (\rho_s - \rho_g)^2}{225}\right]^{1/3} D_p \text{ for } 0.4 < \text{Re} < 500 \qquad (5)$$

where u_t is the terminal velocity, ft/sec.;

$$\text{Re} = \frac{D_p \ \rho_g \ u_t}{\mu}$$ is the Reynolds number, dimensionless.

Kunii and Levenspiel (8) calculated the ratio of terminal

velocity u_t to minimum velocity u_{mf} directly from the equations of minimum fluidization and terminal velocities. For a typical fluidization process of fine solids this ratio is:

$$\frac{u_t}{u_{mf}} = 91.6 \qquad (6)$$

This ratio will govern the range of velocities that will be considered feasible in the analysis of the economic cost model (17), (12), and (10).

Drying Operation

Drying,,as defined by Perry and Chilton (15), generally refers to the removal of a liquid from a solid by evaporation.

In their excellent survey of drying operations in the chemical industry, Perry and Chilton (15) presented a classification of dryers based on heat transfer which is shown in Figure 2 .

As can be seen in Figure 2 , the fluidized-bed drying process is a direct contact continuous drying operation. This process is described by Coulson and Richardson (3) as follows: "Heated air, or hot gas from a burner, is passed via a plenum chamber and a diffuser plate into the fluidized-bed of material, from which it passes to a dust separator. The diffuser plate is fitted with suitable nozzles to prevent back-flow of solids. Wet material is fed continuously into the bed through a rotary valve and mixes immediately with the dry charge. Dry material overflows via a downcomer to an integral after-

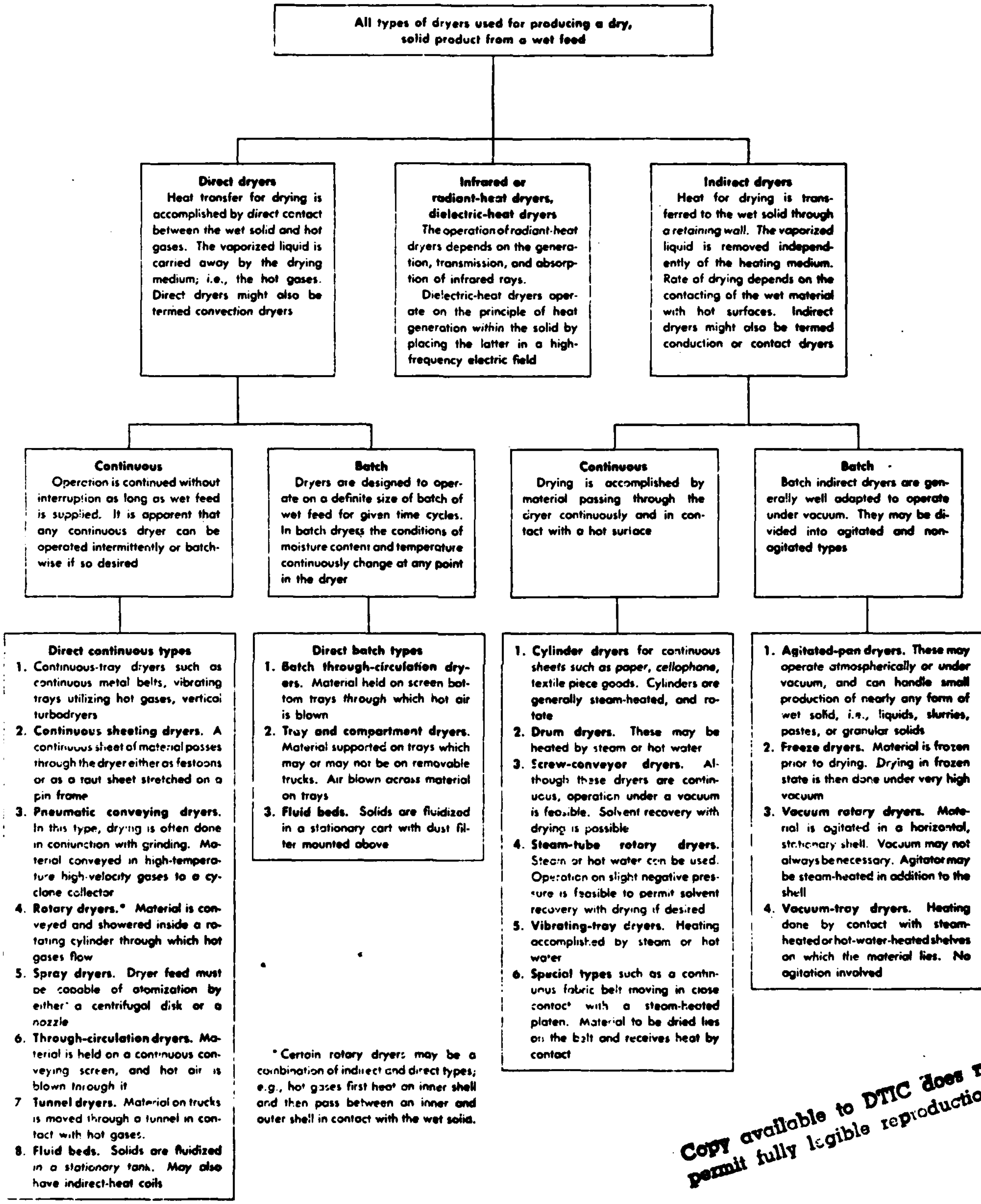

Figure 2 Classification of dryers based on heat transfer. Perry and Chilton (15)

cooler". Figure 3 shows a schematic flow sheet for a typical equipment configuration for a fluidized bed drying opertion.

Since the fluidized-bed drying operation is mainly the removal of moisture content from solid particles by direct contacting with hot air, it requires the basic equipments for heating, blowing of air and fluidizing of solid particles.

As shown in Figure 3 , air enters the heat exchanger with the following specifications:

moisture content, H_i

dry-bulb temperature, T_{Di}

wet-bulb temperature, T_w

Pressure, P_1

The steam enters the heat exchanger with a temperature T_s, and increases the air temperature to T_{go}.

The function of the compressor is to increase the air pressure from P_1 to P_2 to obtain the pressure drop required for fluidization of solid particles. In the fluidization column, the direct contact of heated air with the wet solid results in the removal of solid moisture content from initial quantity X_i to final quantity X_o. As a consequence the moisture content of the air is increased from H_i to H_o.

The design used in this research of the fluidized bed drying process is an approximation of the direct contact rotary dryer. The heat transfer mechanism employed in such a system is given by Perry and Chilton (15) as:

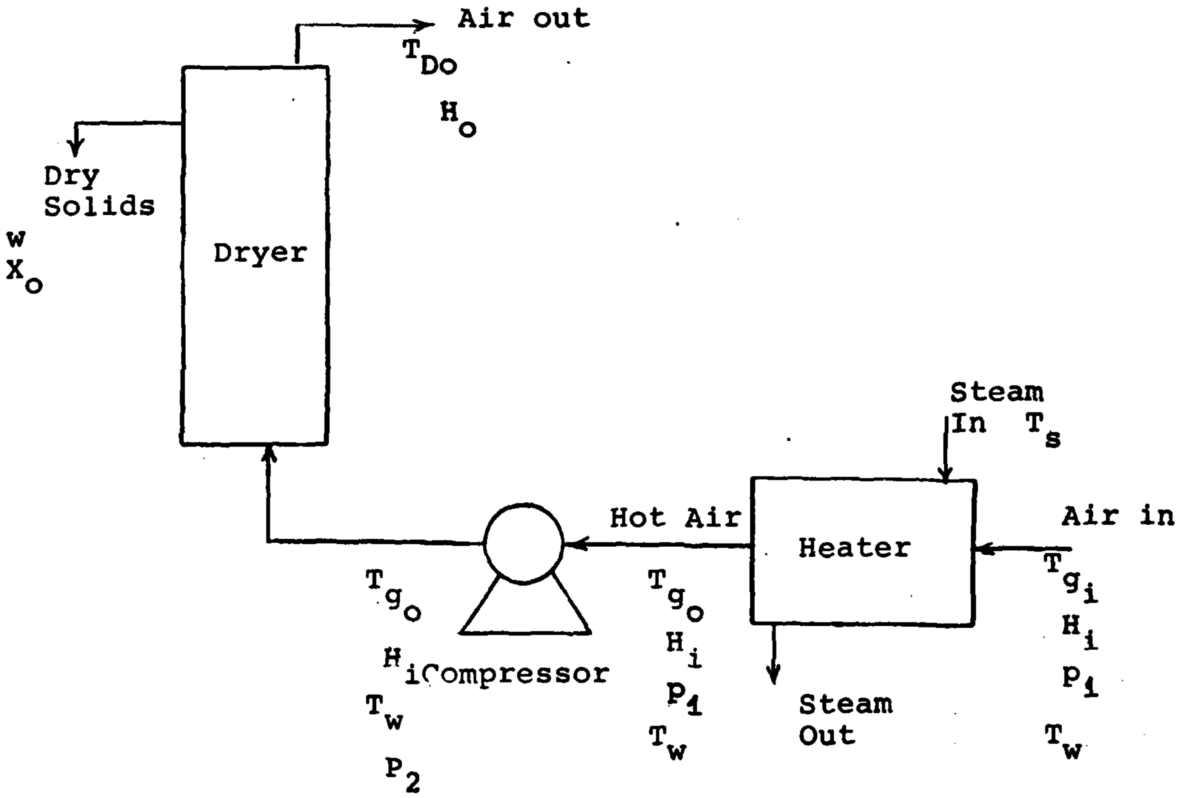

Figure 3 Schematic flow sheet.

$$Qa = U_a \ V \ \Delta T_m \qquad (7)$$

where Qa = total heat transferred, B.t.u/hr.;

U_a = volumetric heat - transfer coefficient, B.t.u./(hr) (cu. ft dryer volume) ($^{\circ}F$);

V = dryer volume, cu. ft.;

ΔT_m = true mean temperature difference between the hot gases and material, $^{\circ}F$.

When a considerable quantity of surface moisture is removed from the solids and the solids temperatures are unknown, a good approximation of (ΔT_m) is the logarithmic mean between the wet-bulb depression of the drying air at the inlet and exit of the dryer.

Data for evaluating U_a were shown in Perry and Chilton (15) and take the form

$$U_a = K \ G^n / D \qquad (8)$$

where K = a proportionality constant;

G = gas mass velocity, lb/(hr.)(sq.ft.) dryer cross-sectional area.

D = dryer diameter, ft;

n = a constant

McCormick, as reported by Perry and Chilton (14) compared all available data and concluded the value of $n = 0.67$ is probably the most truly representative of commercial equipment.

The drying process is related to psychrometry which is concerned with determination of the properties of gas-vapor mixture. The air-water vapor system is by far the one most

commonly encountered. Terminology and relationships pertinent to psychrometry as reported by Perry and Chilton (15) are:

Absolute humidity H equals the pounds of water vapor carried by one pound of dry air.

Percentage absolute humidity is defined as the ratio of absolute humidity to saturation humidity.

Percentage relative humidity is defined as the partial pressure of water vapor in air divided by the vapor pressure of water at the given temperature.

Dew point or saturation temperature is the temperature at which a given mixture of water vapor and air is saturated

Humid heat, C_s is the heat capacity of one pound of dry air and the moisture it contains,

$$C_s = 0.24 + 0.45 H$$

Wet-bulb temperature: is the dynamic equilibrium temperature attained by a water surface when the rate of heat transfer to the surface by convection equals the rate of mass transfer away from the surface.

Psychrometric charts for air-water vapor systems have been developed and are used to find the properties of a system at given dry-bulb and wet-bulb temperature.

Equation (7) will be used for calculation of the dryer volume and equation (8) will be used for determination of dryer diameter.

Compression of Air

Pressure drop is required for the flow of air through the beds of granular solids. It varies due to the variety of granular materials and their packing arrangement. This pressure drop can be created in the fluidized-bed drying system by either compressors or fans. Fans are used for low pressure, in general, for pressure drop of less than 0.5 lb_f/sq. in. (15). Centrifugal compressors are widely used to handle large volumes of gas at pressure rises from 0.5 up to several hundred lb_f/sq. in. The centrifugal compressor cost estimation by Happel and Jordon (7) is given by equation (48) and for fans are

$$\text{Purchased cost} = 6.7\ (\text{cfm})^{0.68}\ \$$$

$$\text{Installed cost} = 2.78\ (6.8)\ (\text{cfm})^{0.68}\ \$$$

Where cfm is the cubic feet per minute gas flow rate comparing this fan cost estimation equations with the centrifugal compressor cost estimation in equation (48) reveals that compressors will be more expensive than fans. Due to the limitation of fans in a fine particles fluidized-bed this study will use the centrifugal compressor.

In any continuous compression process the relation of absolute pressure P to volume V is expressed by the formula

$$PV^n = K \tag{9}$$

The plot of pressure vs. volume for each of exponent n is known as a polytropic curve. Since the work W_p performed in proceeding from P_1 to P_2 along any polytropic curve as in Figure 4 is:

$$W_p = \int_1^2 P\,dv \tag{10}$$

it follows that the amount of work required is dependent upon the polytropic curve involved and increases with increasing values of n. The path requiring the least amount of input work is n = 1, which is equivalent to isothermal compression. For adiabatic compression, i.e. no heat added or taken away during the process, n = k = ratio of specific heat at constant pressure to that at constant volume.

The adiabatic head is calculated by:

$$H_{ad} = \frac{k}{k-1} R\ T_1 \left[\left(\frac{P_2}{P_1}\right)^{\frac{k-1}{k}} - 1 \right] \tag{11}$$

Where R = gas constant = 1544 ft./(lb. mole) ($^{\circ}$K);

T = absolute temperature, $^{\circ}$R.

Since the work performed on the gas is equal to the product of the head and weight of gas handled, the adiabatic horsepower is

$$HP = \frac{WH_{ad}}{550} = \frac{k}{k-1} \frac{WRT_1}{550} \left[\left(\frac{P2}{P1}\right)^{\frac{k-1}{k}} - 1 \right] \tag{12}$$

or

$$HP = 0.00436\ Q\ P_1 \left(\frac{k}{k-1}\right) \left[\left(\frac{P2}{P1}\right)^{\frac{k-1}{k}} - 1 \right] \tag{13}$$

where w = weight of gas, lb./sec.;

H_{ad} = adiabatic head, ft.;

Q = flow rate cu. ft./min.;

P_1 = pressure, lb_f./sq.in.

Equation (13) will be used for compressor capital cost as well as for power consumption cost calculations in developing

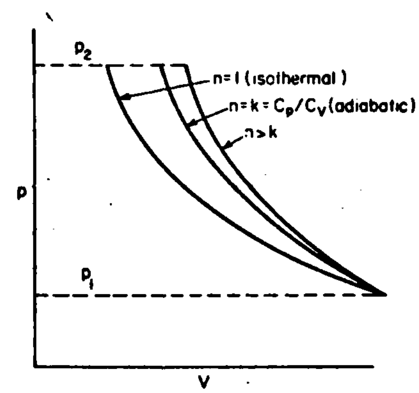

Figure 4 Polytropic compression curve
Perry & Chilton (15)

cost model.

Heat Exchange

The basic design equation for the heat exchanger stated by Perry and Chilton (15) is:

$$dA_H = \frac{dQ_H}{U_H\ T_H} \tag{14}$$

Where dA_H is the element of surface area required to transfer an amount of heat dQ at a point in the exchanger where the over-all bulk temperature difference between the two streams is T_H.

Equation 14 can be integrated to give the outside area required to transfer the total heat load Q_H.

$$A_H = \int_0^Q \frac{dQ_H}{U_H} \tag{15}$$

The heat exchanger used in the design of drying process uses the steam to heat the entering air. If we neglect the heat losses to the surroundings, an energy balance about the heat exchanger gives

$$\binom{\text{heat gained}}{\text{by air}} = \binom{\text{heat lost}}{\text{by steam}} \tag{16}$$

For conditions normally met in practice these continuous operations can be considered equilibrium processes, with exiting gas, and exiting steam all at the same temperature. Thus with F (lb. dry gas/hr) as the feed rate of air, the energy balance becomes

$$F\ C_s\ (T_s - T_{gi}) = U_H A_H\ \Delta T_H \tag{17}$$

Where T_s is the temperature of the steam

T_{gi} is the temperature of the air

C_s is the humid heat capacity of 1 lb. of dry air and the moisture it contains.

For most engineering calculations,

$$C_s = 0.24 + 0.45H$$

Where 0.24 and 0.45 are the heat capacities of dry air and water vapor, respectively and both are assumed constant (15).

Equation (14) is used in estimating the capital cost of the heat exchanger and equation (17) will be used in determining the cost of the steam in developing the cost model.

B. Economical Aspects

The principal economic aim of any company as stated by Freemantle (6) is the efficient use of available resources in order to maximize benefit over the agreed time span. Happel and Jordan (7) discussed in detail the chemical process economics which provide a working tool to assist in applying technical information to the economic design and operation of chemical process plants.

The essential economic relationship of interest is the equipment cost - size relationships which relates the purchased cost of equipment to the size, volume, or power of the equipment.

The general form of the equipment cost - size

relationship is

$$\text{Cost} = k\,(A)^{n1} \tag{18}$$

or

$$\frac{(\text{Cost})_{Lr}}{(\text{Cost})_S} = \left(\frac{A_{Lr}}{A_S}\right)^{n1} \tag{19}$$

where A = some characteristic size measurement such as volume, area, or horsepower.

k = a constant, the value of the cost when A is unity.

Lr and S = subscripts designating larger and smaller sizes.

Extensive studies of process equipment exist which generally report a finding of 0.6 or 0.7 for the value of $n1$ (1). The evidence is so favorable to this value that the so called "point 6 value" has emerged among process engineers to describe the cost - size relationship. That is, in general one expects a 10 percent increase in equipment size to yield a 6 per cent increase in capital costs. The economic impact of an $n1$ value less than unity is that there are economy of large size operation available for many engineering processes.

The equipment capital cost can be transformed to an annual capital - related fixed charge by use of a fixed charge factor r which is defined by Happel and Jordan (7) as:

$$r = m + \frac{1}{n_2} + \frac{i_m - 0.35\,i}{1 - t} \tag{20}$$

where i_m = minimum acceptable return rate on invested capital, (frac./yr.);

$n2$ = allowable life for depreciation of equipment;

$\frac{1}{n2}$ = straight-line depreciation factor;

i = average rate of return on invested capital, (frac./yr.);

t = income tax rate, Federal plus state, (frac./yr.);

m = maintenance charge, (frac./yr.);

While the fixed factor charge will vary equipment and firms, Happel and Jordan argue that a value of $r = 0.5$ is a reasonable approximation for most engineering applications.

III. ECONOMIC MODEL

There are two fundamental economic engineering problems. One consists in simply finding the optimum capital cost for the equipment to accomplish the desired function. Such a situation exists for example in choosing the optimum diameter for tanks and vessels. The other consists in balancing incremental investment costs against possible operating savings. In fluid flow applications for example, one might balance the cost of larger diameter pipe against the saving in pumping costs, or in problems in heat transfer one may balance the cost of extra exchanger surface against the cost of the possible heat saving.

The research work focuses on the optimum capital cost for the equipment of a continuous direct contact drying process as one of the physical applications of the fluidized bed processes.

In studying the economics of process equipment one usually assumes a constant value for all the engineering variables involved except one. The total cost function is then set up in terms of this variable and employs one of the optimization techniques to find an optimum value. The optimization technique that will be used in this study is the Fibonacci search method (9).

The main feature of this model is that according to the calculation of the dryer heat load from the solids feed rate and the moisure to be evaporated, the gas flow

rate which is a function of the gas velocity is calculated. From the known dryer heat load, head - transfer coefficient, and the logarithmic mean temperature difference, the volume of the dryer can be calculated. Relating the crossectional area of the dryer and the height to the gas velocity will develop the economics of the model in terms of gas velocity.

A. Continuous Direct Contact Drying

As discussed in Chapter II, drying is accomplished by the transfer of sensible heat from a gas to a solid. The evaporation of liquid from the solid and removal of the evaporated liquid as a vapor in the exit gas. Thus, a dryer is both a heat - transfer and a mass - transfer device (8).

The total cost of the drying operation consists of the capital cost of the dryer, the air heater, and compressor and the operating cost of steam and electric power consumption. We will discuss each separately and then sum all these items to get the total cost model.

1. The Dryer Capital Cost

An overall humidity or water balance gives:

$$W\ (X_i - X_o) = F\ (H_o - H_i) \qquad (21)$$

where W = pounds of dry solids to be treated per hour;

X_o = leaving solid moisure content (lb. water/lb. dry solid);

X_i = entering solid moisure content (lb. water/lb. dry solid);

H_o = humidity, or moisture content, of leaving gas (lb. water/lb. dry gas);

H_i = humidity, or moisture content, of entering gas (lb. water/lb. dry gas);

F = lb. dry gas / hr.

Thus,

$$F = \frac{W(X_i - X_0)}{(H_o - H_i)}$$

But since F (lb dry gas/hr) = A_c (cross sectional area sq. ft. x ρ (density of gas $\frac{\text{lb.}}{\text{cu.ft.}}$) x u (velocity $\frac{\text{ft.}}{\text{hr.}}$)

Then
$$F = \frac{(X_i - X_o)}{(X_o - H_i)} = A_c \rho u \tag{22}$$

Since G, the mass rate of flow of the gas in the dryer lb/(hr.) (sq. ft.) = ρ u from Equation (22) we have

$$F = A_c \rho u = A_c G$$

i.e.

$$A_c = \frac{F}{G}$$

But A_c (cross sectional area = $\frac{\pi}{4} D^2$

Where D is the diameter of the dryer

Thus
$$D = \left[\frac{4}{\pi} \frac{F}{G}\right]^{0.5} \tag{23}$$

or
$$D = \left[\frac{4}{\pi} \frac{W(X_i - X_o)}{\rho W (H_o - H_i)}\right]^{0.5} \tag{24}$$

The rate of heat transfer in the dryer is given by:

$$Qa = U_a V \Delta T_m \tag{25}$$

Where Qa = the rate of heat transfer in the dryer (Btu/hr)

U_a = the overall volumetric heat - transfer coefficient

between gas and solid (Btu/(hr.) (cu.ft.) ($^{\circ}$F.))

V = the volume of the dryer (cu.ft.)

ΔT_m = overall logarithmic temperature difference between gas and solid ($^{\circ}$F.)

$$\Delta T_m = \frac{(T_{Di} - T_w) - (T_{Do} - T_w)}{\text{Ln}\ \dfrac{T_{Di} - T_w}{T_{Do} - T_w}} \tag{26}$$

Where T_{Di} = inlet gas dry bulb temperature ($^{\circ}$F.)

T_{Do} = outlet gas dry bulb temperature ($^{\circ}$F.)

T_w = wet bulb gas temperature ($^{\circ}$F.)

Assuming the latent heat of vaporization as 1,000 Btu/lb, this gives

$$Qa = W(X_i - X_o)\ (1{,}000)\ \text{Btu./ hr.} \tag{27}$$

The work of McCormick (11) gives an equation for U_a:

$$U_a = 20\ G^{0.67}/D \tag{28}$$

Substituting for D from equation (23) gives

$$U_a = 20\ G^{0.67} \ /\ (\tfrac{4}{\pi}\ \tfrac{F}{G})^{0.5}$$

$$= \frac{20(\pi)^{0.5}\ G^{1.26}}{2\ \ F^{0.5}} \qquad = 17.725\ \frac{G^{1.26}}{F^{0.5}}$$

$$= 17.725\ \frac{(\rho\ u)^{1.26}}{F^{0.5}} \tag{29}$$

The volume of the dryer may be calculated from Equation (25)

$$V = \frac{Qa}{U_a\ \Delta T_m}$$

Substituting for Qa from Equation (27) and for U_a from Equation (29) we get

$$V = \frac{1000\ W\ (X_i - X_o)\ F^{0.5}}{17.725\ (\rho\ u)^{1.26}\ \Delta T_m}$$

$$= \frac{56.417\ W\ (X_i - X_o)\ F^{0.5}}{(\rho\ u)^{1.26}\ \Delta T_m} \qquad (30)$$

Since $F = \frac{W(X_i - X_o)}{H_o - H_i}$

then

$$V = \frac{56.417\ W\ (X_i - X_o) \left[\frac{W(X_i - X_o)}{H_o - H_i}\right]^{0.5}}{(\rho\ u)^{1.26}\ \Delta T_m}$$

$$= \frac{56.417\ \ W(X_i - X_o)^{1.5}}{(X_o - H_i)^{0.5}\ (\rho\ u)^{1.26}\ \Delta T_m} \qquad (31)$$

The installed cost of the dryer may be expressed as a function of its peripheral area, $A_p = \pi\ D\ L$,

where L is the length of the dryer expressed in feet.

Since $V = \frac{\pi}{4} D^2 L$

$$L = \frac{4\ V}{\pi\ D^2}$$

and $A_p = \pi\ D\ \frac{4}{\pi}\ \frac{V}{D^2}$

$$= \frac{4\ V}{D} \qquad (32)$$

Substituting the expressions for V and D from Equations (24) and (31) in the above equation for A_p gives

$$A_p = 4\ \frac{56.417 \left[W(X_i - X_0)\right]^{1.5}}{(H_o - H_i)^{0.5}\ (\rho\ u)^{1.26}\ \Delta T_m} \left[\frac{\pi\ \rho\ u\ (H_o - H_i)}{4\ W(X_i - X_o)}\right]^{0.5}$$

$$= 200.0 \frac{W\ (X_i - X_o)}{(\rho\ u)^{0.76}\ \Delta T_m} \tag{33}$$

From the correlations of the costs of dryers as presented by Happel and Jordan (15), it can be found that the installed cost of a carbon steel dryer is

$$\text{Purchased cost} = 195\ (A_p)^{0.8} \quad \$$$

$$\text{Installed cost} = 3.0\ (195)\ A_p^{0.8} \quad \$ \tag{34}$$

Where A_p is the peripheral area of the dryer

The installed cost of the dryer is then

$$\text{Cost} = 585 \left[\frac{200.0\ W\ (X_i - X_o)}{(\rho\ u)^{0.76}\ \Delta T_m} \right]^{0.8} \tag{35}$$

The annual fixed charges on the dryer is

$$\text{Cost} = r \left\{ 585 \left[\frac{200\ W\ (X_i - X_o)}{(\rho\ u)^{0.76}\ \Delta T_m} \right]^{0.8} \right\} \tag{36}$$

where r is annual fixed charge factor

2. The Air Heater Capital Cost

The heater required for the heating of the inlet air may be assumed to be a standard steam - heated exchanger of the usual shell and tube type, and made of carbon steel.

From the energy balance of equation (21) the heating load will be

$$Q_H = \frac{W(X_i - X_o)}{H_o - H_i} \left[0.24\ (T_{go} - T_{gi}) + 0.45\ H_i (T_{go} - T_{gi}) \right] \tag{37}$$

where T_{go} = outlet gas temperature (°F)

T_{gi} = inlet gas temperature (°F)

$$Q_H = U_H A_H \Delta T_H \tag{38}$$

where U_H = the overall heat - transfer coefficient Btu/(hr.)(sq. ft.)(^{o}F.)

$$\Delta T_H = \frac{(T_s - T_{gi}) - (T_s - T_{go})}{\mathrm{Ln}\,\dfrac{(T_s - T_{gi}}{T_x - T_{go}}}$$ ^{o}F., the mean logarithm temperature difference between the air and steam

A_H = the peripheral area of the heater (sq. ft.) from Equations (37) and (38)

$$A_H = \frac{W(X_i - X_o)}{(H_o - H_i)\ U_H\ \Delta T_H}\left[0.24(T_{go} - T_{gi}) + 0.45\ H_i\ (T_{go} - T_{gi})\right] \tag{39}$$

The purchased and installed costs of heat exchangers have been estimated by Happel and Jordan (15) as:

Purchased cost = $0.5\ (A_H)^{0.62}$ \$

Installed cost = $346\ (A_H)^{0.62}$ \$

Thus the annual fixed charge on the air heater will be

$$r\ \left\{346\left(\frac{W(X_i - X_o)}{(H_o - H_i)\ U_H\ \Delta\ T_H}\left[0.24\ (T_{go} - T_{gi}) + 0.45\ H_i\ (T_{go} - T_{gi})\right]\right)\right\}\ \$/\mathrm{yr.} \tag{40}$$

3. The Steam Cost

The cost of the steam used to heat the air before it enters the dryer, assuming 24 $\frac{\text{hr.}}{\text{day}}$ x 330 $\frac{\text{day}}{\text{year}}$ ≃ 8000 hours of operation/year will be

$$\text{Cost} = 8{,}000\ C_s \left[\frac{W(X_i - X_o)}{H_o - H_i}\right] \left[0.24\ (T_{go} - T_{gi}) + 0.5\ H_i\ (T_{go} - T_{gi})\right] \tag{41}$$

Where C_s = cost of steam $/$10^6$ Btu. Happel and Jordan (7) estimated C_s to be 0.8 $/$10^6$ Btu. in 1975 and we will assume the energy cost is increased three times.

The gases leaving the dryer will be at a fairly low temperature and will be assumed to have no value.

4. The Compressor Capital Cost

Power consumption is an important factor in any process using fluidized beds. Figure (5) shows the power required to supply sufficient incoming fluidizing gas at pressure P_3 is that necessary to compress that quantity of gas from pressure P_1 to P_2.

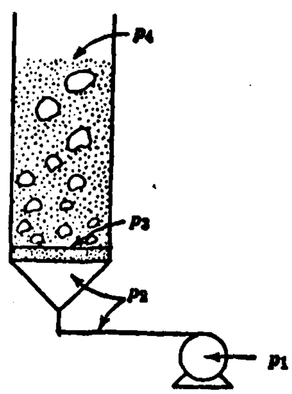

Figure 5. Pressure at various locations in fluidized bed.

This is found by taking a mechanical energy balance across the compressor. Thus the shaft work required per mole of gas compressed, as stated by Kunii and Levenspiel (8) is:

$$- W_s = \frac{g}{g_c} \Delta L + \int_{P_1}^{P_2} \frac{dp}{\rho} + \frac{u_1^2 - u_1^2}{2 g_c} + \text{(friction loss)} \quad (42)$$

For adiabatic reversible operations with negligible kinetic and potential energy effects, this reduces to the ideal work.

The adiabatic head H_{ad} is calculated as

$$H_{ad} = \frac{k}{k-1} RT_1 \left[\left(\frac{P_2}{P_1}\right)^{\frac{k-1}{k}} - 1 \right] \quad (43)$$

where k = specific heat ration = 1.4

R = gas constant (1544 ft/(lb. mole) ($^{\circ}$R)

T = absolute temperature , $^{\circ}$R.

Since the work performed on the gas is equal to product of the head and the weight of gas handled, the adiabatic horsepower is: (11)

$$HP = \frac{W H_{ad}}{550} = \frac{k}{k-1} \frac{W R T_1}{550} \left[\left(\frac{P_2}{P_1}\right)^{\frac{k-1}{k}} - 1 \right] \quad (44)$$

or

$$HP = 0.2616\ Q\ P_1 \left(\frac{k}{k-1}\right) \left[\left(\frac{P_2}{P_1}\right)^{\frac{k-1}{k}} - 1 \right] \quad (45)$$

where W = weight of gas lb/sec.;

H_{ad} = adiabatic head, ft;

Q = flow rate, cu. ft/hr.;

P = pressure lb_f./sq. in.

The flow rate Q is related to the velocity u by the relation

$$Q = \frac{F}{\rho} = \frac{W(X_i - X_o)}{\rho\ (H_o - H_i)} \tag{46}$$

thus

$$HP = 0.2616\ \frac{W(X_i - H_o)}{\rho\ (H_o - H_i)}\ P_1 \left(\frac{k}{k-1}\right) \left(\frac{P_2}{P_1}\right)^{\frac{k-1}{k}} - 1 \tag{47}$$

The purchased and installed cost of the compressor as calculated by Happel and Jordan (7) are:

$$\text{Purchased cost} = 645\ (HP)^{0.8}\ \$ \tag{48}$$

$$\text{Installed cost} = 3.1\ (645)\ (HP)^{0.8}\ \$$$

Substituting in equation (47) gives:

$$\text{Installed cost} = 3.1\ (645)\ \left[0.2616\ \frac{W(X_i - X_o)}{\rho\ (H_o - H_i)}\ P_1 \left(\frac{k}{k-1}\right) \left(\frac{P_2}{P_1}\right)^{\frac{k-1}{k}} - 1\right]^{0.8}$$

$$= 684.0\ \left[\frac{W(X_i - X_o)}{\rho\ (H_o - H_i)}\ P_1\ \left(\frac{k}{k-1}\right) \left(\frac{P_2}{P_1}\right)^{\frac{k-1}{k}} - 1\right]^{0.8} \tag{49}$$

Assuming that the effect of the distributer of the gas in the fluidized column is negligible

$$\Delta P = P_3 - P_4 \simeq P_2 - P_1$$

but ΔP is calculated for the fluidized bed as:

$$\Delta P = \frac{L}{144}\ (1-\varepsilon)\ (\rho_s - \rho_g)\ \frac{g}{g_c}\ (\text{lb}_f\text{./sq. in.}) \tag{50}$$

where L = bed height, ft.;

ρ = gas density, lb./cu.ft.;

ρ_s = solid density, lb./cu.ft.;

ε = bed voidage fraction, dimensionless;

g = acceleration of gravity, 32.2 ft./(sec.)(sec.);

g_c = conversion factor, lb. ft./(lb_f.)(sec.)(sec.).

P_1 is usually the atmospheric pressure = 14.7 lb_f./sq. in.

Thus $\Delta P = \frac{L}{144}(1-\varepsilon)(\rho_s - \rho_g)\frac{g}{g_c} \simeq P_2 - P_1$

$$\frac{L}{144}(1-\varepsilon)(\rho_s - \rho_g)\frac{g}{g_c} \simeq P_2 - 14.7$$

$$P_2 = 14.7 + \frac{L}{144}(1-\varepsilon)(\rho_s - \rho_g)\frac{g}{g_c} \qquad (51)$$

Since $L = \frac{4\ V}{\pi\ D^2}$, substituting for V and D from

Equations (24) and (31), we get

$$L = \frac{4\ (56.417)\ W\ (X_1 - X_o)\ F^{0.5}}{\pi\ (\rho\ u)^{0.66}\ \Delta T_m\ \left(\frac{4}{\pi}\frac{F}{\rho\ u}\right)}$$

i.e

$$L = \frac{56.417\ W\ (X_i - X_o)\ (\rho\ u)^{0.34}}{\Delta T_m\ \frac{W\ (X_i - X_o)}{(H_o - H_i)}}$$

$$L = \frac{56.417\ (\rho\ U)^{0.34}\ (H_o - H_i)}{\Delta T_m} \qquad (52)$$

Equation (51) becomes

$$P_2 = 14.7 + 0.4\ \frac{(\rho\ u)^{0.34}\ (1-\varepsilon)\ (\rho_s - \rho)}{\Delta T_m}\ \frac{g}{g_c} \qquad (53)$$

Substitute for P_2 in Equation (29)

$$\text{Installed cost} = 5873.68 \left\{ \frac{W\ (X_i - X_o)}{\rho\ (H_o - H_i)} \left(\frac{k}{k-1}\right) \left[\left(\frac{P_2 \text{ expression}}{14.7}\right)^{\frac{k-1}{k}} - 1 \right] \right\}^{0.8} \ \$ \qquad (54)$$

The annual fixed cost of the compressor is

$$r = 5873.68 \left\{ \frac{W(X_i - X_o)}{\rho\ (H_o - H_i)} \left(\frac{k}{k-1}\right) \left[\left(\frac{P_2 \text{ expression}}{14.7}\right)^{\frac{k-1}{k}} - 1 \right] \right\}^{0.8} \ \$/\text{yr.} \qquad (55)$$

5. Compressor Power Consumption Cost

The cost of power consumption is calculated as follows:

1. The annual hours = 8,000 hr.
2. The horsepower of the compressor = HP from Equation (57)
3. Efficiency is taken to be 0.75 (8)
4. The cost of electrical power Ce is taken as \$0.24/kw.hr. since it was estimated in 1975 by Happel and Jordon (7) as \$0.08/Kw/hr.

The power consumption cost = $\frac{8{,}000\ (HP)\ (Ce)}{0.75}$ \$

$$= 41018.88\ Ce\ \frac{W(X_i - X_o)}{\rho\ (H_o - H_i)}\left(\frac{k}{k-1}\right)\left[\left(\frac{P_2\ \text{expression}}{14.7}\right)^{\frac{k-1}{k}} - 1\right] \quad (56)$$

B. Total Cost

The total capital cost is obtained by adding the cost of each equipment, represented by Equations (36), (40) and (55).

$$\text{Capital Cost} = r\left\{585\left[\frac{200\ W\ (X_i - X_o)}{(\rho\ u)^{0.76}\ \Delta\ T_m}\right]^{0.8}\right.$$

annual fixed charge on the dryer

$$+\ r\left\{346\left(\frac{W(X_i - X_o)}{(H_o - H_i)\ U_H\ \Delta T_H}\left[0.24\ (T_{go} - T_{gi}) + 0.45\ H_i\ (T_{go} - T_{gi})\right]\right)^{0.62}\right\}$$

annual fixed charge on the heater

$$+\ r\left\{5873.68\left(\frac{W(X_i - X_o)}{\rho\ (H_o - H_i)}\ \frac{k}{(k-1)}\left[\left(\frac{P_2\ \text{expression}}{14.7}\right)^{\frac{k-1}{k}} - 1\right]\right)^{0.8}\right\}$$

annual fixed cost of compressor (57)

The total operating cost is considered as the cost of the steam used to heat the air before entering the dryer, and the electrical power consumption by the compressor.

$$\text{Operating Cost} = 8{,}000\ C_s \Big[\frac{W\ (X_i - X_o)}{(H_o - H_i)}\ 0.24\ (T_{go} - T_{gi}) +$$

$$0.45\ H_i\ (T_{go} - T_{gi}) \Big]$$

$$+ 41018.88\ Ce\ \frac{W\ (X_i - X_o)}{\rho\ (H_o - H_i)}\ \left(\frac{k}{k-1}\right)$$

$$\left[\left(\frac{P_2\ \text{expression}}{14.7} \right)^{\frac{k-1}{k}} - 1 \right] \tag{58}$$

Thus the total cost = Capital cost + Operating cost or

$$\text{Total Cost} = r \left\{ 585 \left[200\ \frac{W\ (X_i - X_o)}{(\rho\ u)^{0.76}\ \Delta T_m} \right]^{0.8} \right\}$$

$$+ r \left\{ 346 \left(\frac{W\ (X_i - X_o)}{(H_o - H_i)\ U_H\ \Delta T_H} \Big[0.24\ (T_{go} - T_{gi}) + \right.\right.$$

$$\left.\left. 0.5\ H_i\ (T_{go} - T_{gi}) \Big] \right)^{0.62} \right\}$$

$$+ r \left\{ 5873.68 \left(\frac{W\ (X_i - X_o)}{\rho\ (H_o - H_i)}\ \left(\frac{k}{k-1}\right) \right.\right.$$

$$\left.\left. \left[\left(\frac{P_2\ \text{expression}}{14.7} \right)^{\frac{k-1}{k}} - 1 \right] \right)^{0.8} \right\}$$

$$+ 8{,}000\ C_s \left(\frac{W\ (X_i - X_o)}{(H_o - H_i)} \Big[0.24\ (T_{go} - T_{gi}) + \right.$$

$$\left. 0.45\ H_i\ (T_{go} - T_{gi}) \Big] \right)$$

$$+ 41018.88\ Ce\ \frac{W\ (X_i - X_o)}{\rho\ (H_o - H_i)}\ \left(\frac{k}{k-1}\right) \left[\left(\frac{P_2\ \text{expression}}{14.7} \right)^{\frac{k-1}{k}} - 1 \right]$$

where P_2 expression is

$$P_2 = 14.7 + \frac{0.4\ (\rho\ u)^{0.14}}{\Delta T_m}\ (1 - \varepsilon)\ (\rho_s - \rho)$$

The values of the parameters are specified as follows:

r = the annual fixed charge factor = 0.5

W = pounds of dry solids to be treated per hour

X_i = inlet solid moisure content = 10%

$= \frac{0.10}{0.90} = 0.111$ lb. water/lb. dry solid

X_o = outlet moisure content = 1%

$= \frac{0.01}{0.99} = -.0101$ lb. water/lb. dry solid

ρ = air density = 0.074 lb/cu. ft.

$$\Delta T_m = \frac{(T_{Di} - T_W) - (T_{Do} - T_W)}{\mathrm{Ln}\ \frac{T_{Di} - T_w}{T_{Do} - T_W}}$$

T_w = wet - bulb air temperature = 92°F.

T_{Di} = inlet dry - bulb temperature = 200°F.

T_{Do} = outlet dry - bulb temperature is obtained according to the values assumed for H_o (moisture content of leaving air) by using air - water humidity charts.

Table 1 contains the different assumed values that will be used for running the model.

Table I

Values of Outlet Moisture Content of Air

Relative humidity of exit air	H_o $\left(\frac{\text{lb. water}}{\text{lb. dry air}}\right)$	T_{Do} ($^\circ F$)	ΔT_m ($^\circ F$)
90	0.0325	95	29.3
60	0.0300	105	44.8
40	0.0275	115	55.0
20	0.0230	132	69.0
10	0.0200	148	80.0
5	0.0135	174	93.6

T_{go} = outlet air temperature = 200°F.

T_{gi} = inlet air temperature = 70°F.

H_i = moisture content of entering air = 0.0008 $\frac{\text{lb. water}}{\text{lb. dry gas}}$

U_H = the overall heat - transfer coefficient for the heater = 5 Btu/(hr.) (sq. ft.) (°F.)

$$\Delta T_H = \frac{(T_x - T_{gi}) - (T_s - T_{go})}{\text{Ln}\ \frac{T_s - T_{gi}}{T_s - T_{go}}} = \frac{(250-70) - (250-200)}{\text{Ln}\ \frac{250-70}{250-200}} = 102^\circ F.$$

T_s = steam temperature = 250°F.

k = specific heat ratio for air = 1.4

$\left(\frac{k-1}{k}\right) = \left(\frac{1.4-1}{1.4}\right) = .29$

ρ_s = solid density = 100 lb/cu.ft.

ε = bed voidage = 0.40

C_s = cost of steam \$/10^6 B.t.u = 2.4

It is clear that some cost terms are a direct function of the gas velocity u as in case of dryer capital cost, compressor capital cost, and electrical power consumption cost while other cost terms are independent upon gas velocity as the heater capital cost and steam cost. For stipulated drying process conditions, the economical gas velocity is obtained by:

Minimizing: Total cost = G(u)

subject to

$$u_{mf} \leq u \leq u_t$$

where u_{mf} = the minimum fluidization velocity

u_f = the terminal velocity

The optimization technique used in this work is a region elimination method which narrow successively the region contain the optimal solution. Figure 6 shows the relative efficiency of various search optimization methods. Fibonacci search method is used in this work which is the efficient one. The results will be presented as an optimal range rather than a fixed value (9). The Fibonacci algorithm is described in Appendix A.

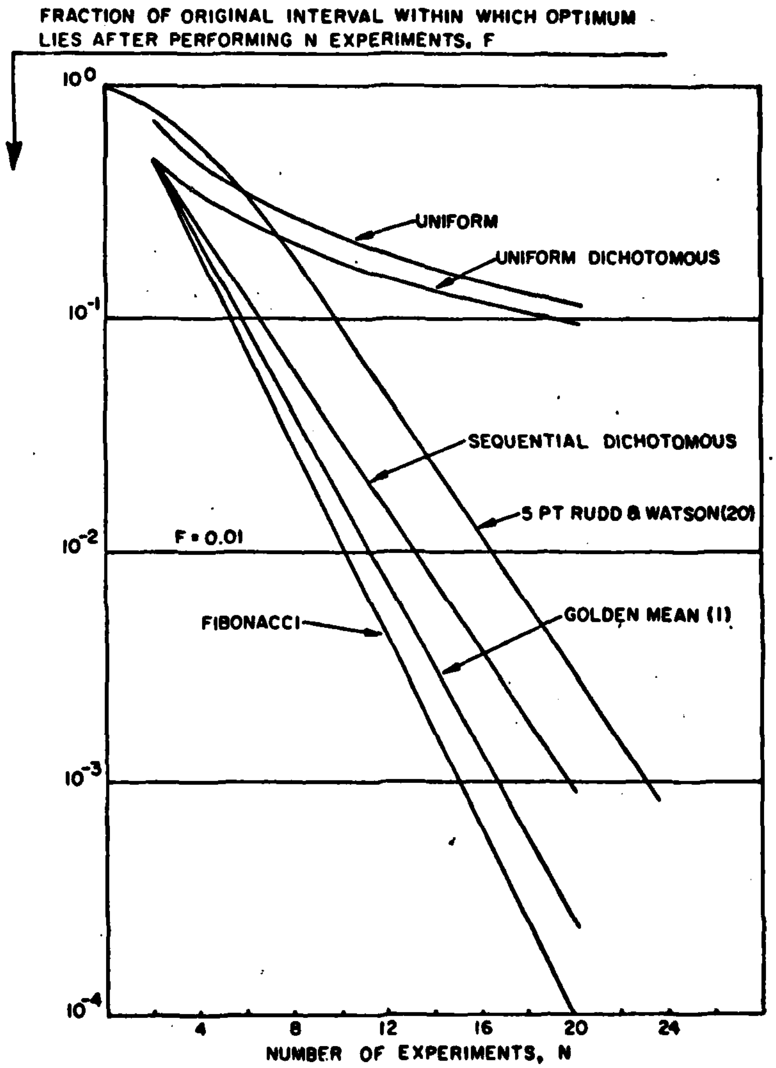

Figure 6 Relative efficiency of various search methods. Happel and Jordan (7)

IV. PRESENTATION AND DISCUSSION OF RESULTS

The model developed in the previous chapter is used to study a fluidized bed drying process. The optimal design of a fluidized bed drying process is affected by a number of factors, most important by the fluidized velocity and the amount of solid to be dried. Each component of the cost model is a function of one or both of these factors, and we are primarily interested in the sentitivity of cost to changes in these factors. Other factors which have a minor effect on costs will also be studied. The analysis will address both the analytical and technical reasons for the cost behavior.

This chapter will be divided into three sections. The first section will analyze the behavior of the total cost. The behavior of component costs will be examined in section two. And finally, the third section will review the general finding discuss their validation, and offer recommendations for further model development.

Since the minimum fluidization process is determined by the specification of the fluidized bed drying operation, a range of this velocity is chosen according to practical application and examples in Leva (11) or Kunii and Levenspiel (8). This range is from 200 (ft/hr) to 1000 (ft/hr) and the base case used for comparisons will be at 600 (ft/hr). The solid flow rate for the base case will be kept at 1000 (lb/hr) and the moisture content will be

at 0.0325 (lb. water/lb. dry air)

Other factors which are held constant in the base case and can also affect capital cost, power cost, or both are as follows:

1. Density of solid material = 100 (lb/cu.ft.)
2. Moisture content of outlet air = 0.0325 $\frac{(\text{lb. water})}{(\text{lb. dry air})}$
3. Electricity cost = .24 ($ /Kw. hr.)
4. Steam cost = 2.4 ($ /10^6 Btu)
5. Annual fixed charge factor = 0.5

The sensitivity of results to these assumptions will be examined in subsequent sections of this chapter.

The relative importance of equipment and power components in the total cost of the drying operation are shown in Table 2. The cost values are reported for three different optimum fluidized velocity levels. The total cost is 1.1 million for each of the dryer designs. The following points emerge with respect to total system cost.

1. The most important element in cost is the compressor. It represents 62.2 - 66.3 per cent of the total cost.
2. The second most important component of the cost is the dryer which represents 32.2 - 36.5 per cent of total cost.
3. Heater costs are independent of fluidized velocity and represent a very small percentage (0.087) of total cost.

4. The three equipment components - compressor, dryer, and heater - account for 98.7 per cent of the total cost while the power consumption cost is 1.3 per cent of the total cost.

Table 2

Percentage of Equipment Capital Cost and Power Consumption Cost

Operating Fluidized Velocity (ft/hr)	Cost Component	Cost Value x 10^3 ($/year)	Percentage from Total Cost (%)
	Dryer	405.1	36.5
	Heater	.9612	0.087
10,800	Compressor	689.6	62.2
	Steam	5.879	0.53
	Electricity	7.798	0.70
	Total Cost	1,109	100.0
	Dryer	379.9	34.3
	Heater	.9612	0.087
12,000	Compressor	713.0	64.4
	Steam	5.879	0.53
	Electricity	8.062	0.73
	Total	1,108	100.0
	Dryer	358.6	32.4
	Heater	.9612	0.087
13,200	Compressor	734.7	66.3
	Steam	5.879	0.53
	Electricity	8.308	0.75
	Total	1,108	100.00

5. For a given dryer efficiency, higher operating velocities call for a substitution of compressor horsepower for dryer peripheral area. As shown in Table 2 an increase in the operating fluidized velocity from 10,800 (ft/hr) to 13,200 (ft/hr) raises the compressor cost while the dryer cost falls. The former occurs because the compressor cost is a direct function of its horsepower which increases with an increase in the pressure required to increase the fluidization velocity. The latter occurs due to the decrease in the required dryer peripheral area. The decrease in peripheral area results from the increase of the air flow rate required to remove a fixed moisture content from a given constant solid material. This means that for a given drying operation, an increase in fluidization velocity will increase the dry efficiency if we hold the dryer peripheral area constant and consequently increase the compressor cost. However, if we want the same drying efficiency with velocity increase a decrease in drying peripheral area will be feasible which means a decrease in the dryer cost.
6. Steam cost, which represents about 42 per cent of the total power consumption cost is independent of fluidized velocity.

7. Electrical power consumption, varies directly with velocity and represent 58 per cent of the total power cost.

A. Capital and Power Cost

Table 3 and Figure 7 illustrate the sensitivity of total cost to variation in velocity. Two points emerge in the analysis of cost and velocity. First, the optimum velocity is independent of the minimum fluidization velocity. As shown in Table 3 minimum fluidization velocity is changed from 200 (ft./hr.) to 1000 (ft./hr.) with the results that the optimal velocity is practically the same and is near the value 12,000 (ft./hr.). Since the optimal velocity is the same the optimal equipment design is the same and minimum system cost is given at about $1100 thousand as shown in Table 3. Figure 7 shows the case of minimum fluidization velocity = 600 (ft./hr.).

The results of Table 3 will hold so long as the operating fluidization range, which is bounded by the minimum fluidization velocity and the terminal fluidization velocity, includes the optimal velocity value. When this is not the case, the minimum cost will occur at the terminal velocity. This fact is shown in Figures 8 and 9 for minimum fluidization velocity values of 30 and 50 (ft./hr.) In the cases optimal velocity is outside the operating velocity constant ratio of the model. Figure 7 shows how cost behave for the base case with minimum fluidization velocity of 600 (ft./hr.).

The optimal velocity is 12,202 (ft./hr.). Cost rise sharply for velocity values below the optimum and rises gradually as velocity is increased by and the optimum range.

The extent of the cost penalty for operation outside the optimum velocity range is illustrated in Table 4. Table 4 shows departure from the optimum velocity, 12,202 (ft./hr.), and the cost penalties which these departures impose on the system. For example, if the system were designed to operate at minimum fluidization velocity if 600 ft./hr., system costs would be $2.6 million or 137.30 per cent more than would be incurred if the system subject to this minimum velocity were designed to operate at the optimum fluidization velocity of 12,202. If the system were designed to one-half the optimum velocity (i.e. velocity of 6,100 (ft./hr.), cost would be $1.16 million or 4.7 per cent greater than the minimum system cost.

A similar cost penalty occurs with systems designed to operate at fluidization velocities in excess of the optimum. For example at twice the velocity level, the system design would cost $1.15 million or 4.15 per cent more than the optimal design. At very high velocities, the cost penalty is not nearly as great as it is for very low fluidization velocities. For example, at the terminal velocity of 54,600 (ft./hr.) the system design would cost 202 thousands ($/yr.) or about 18.23 per cent more than the minimum cost design.

The greater cost penalty of suboptimal velocity designs reflects the fact that the compressor cost is minimized at the minimum fluidization velocity while the dryer cost is minimized at the terminal velocity.

Table 3

Capital and Power Cost

at Different Minimum Fluidization Velocities

Case	Minimum Fluidization Velocity (ft./hr.)	Optimal Velocity Range (ft./hr.)		Minimum Cost Range x 10^3 (\$/yr.)	
		From	To	From	To
1	200	12,231	12,184	1,107.77	1,107.77
2	600	12,202	12,345	1,107.77	1,107.78
3	1000	12,220	11,981	1,107.76	1,107.80

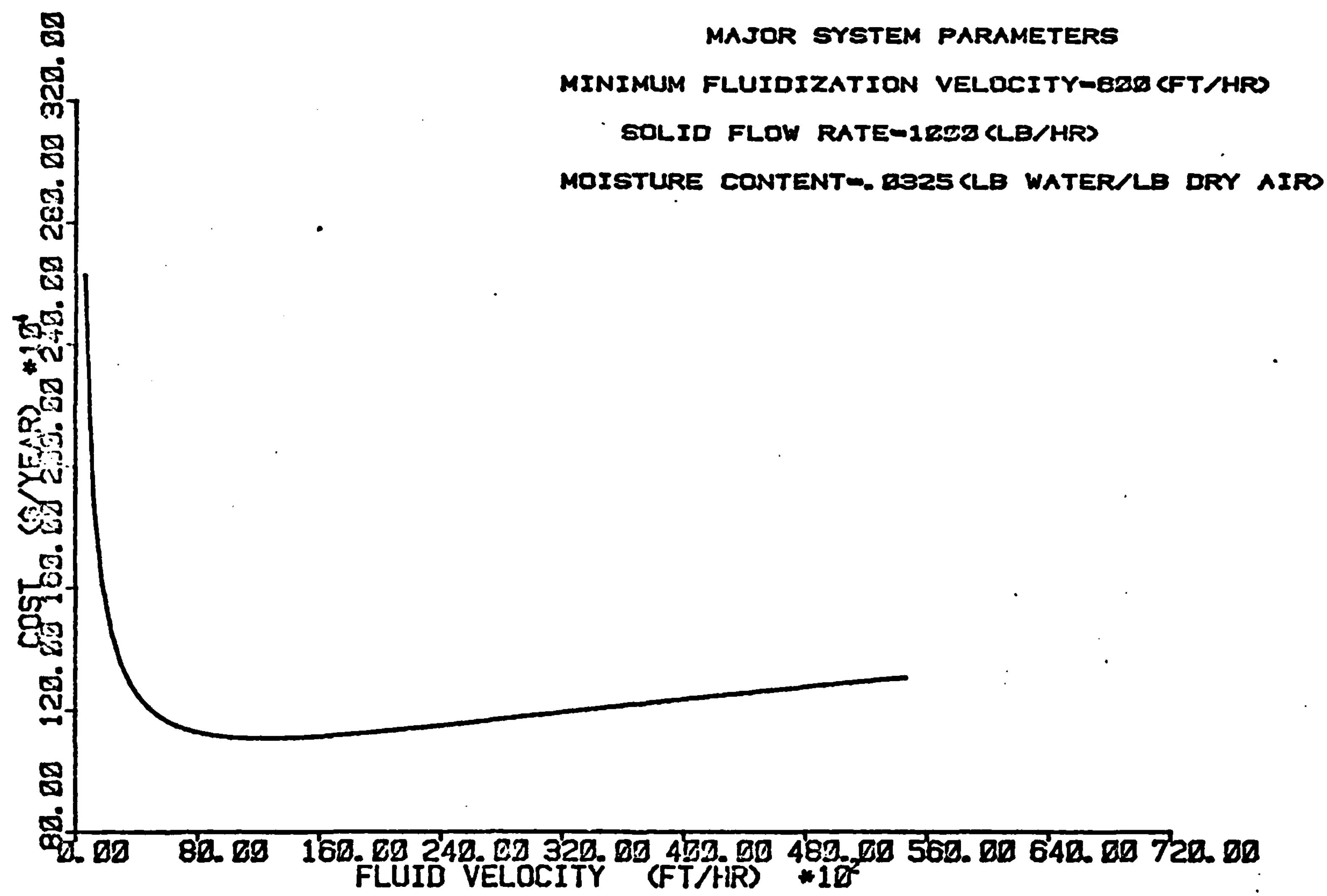

FIGURE (7). CAPITAL AND POWER COST

Table 4

Cost Penalties for Operating

Outside the Optimal Velocity Range (12,202 ft./hr.)

Case	Fluidized Velocity (ft./hr.)	Difference from Optimal Velocity (ft./hr.)	Cost $.10^3$ (\$/yr.)	Difference from Minimum cost x 10^3 (\$/hr.)	Per cent Increase
1	600	-11602	2,629	1,521	137.30
2	6100	-6102	1,160	52	4.7
3	13800	1598	1,109	1	0.09
4	20400	8198	1,133	25	2.26
5	24400	12198	1,154	46	4.15
6	33600	21398	1,203	95	8.57
7	40200	27998	1,239	131	11.82
8	46800	34598	1,272	164	14.80
9	54600	42398	1,310	202	18.23

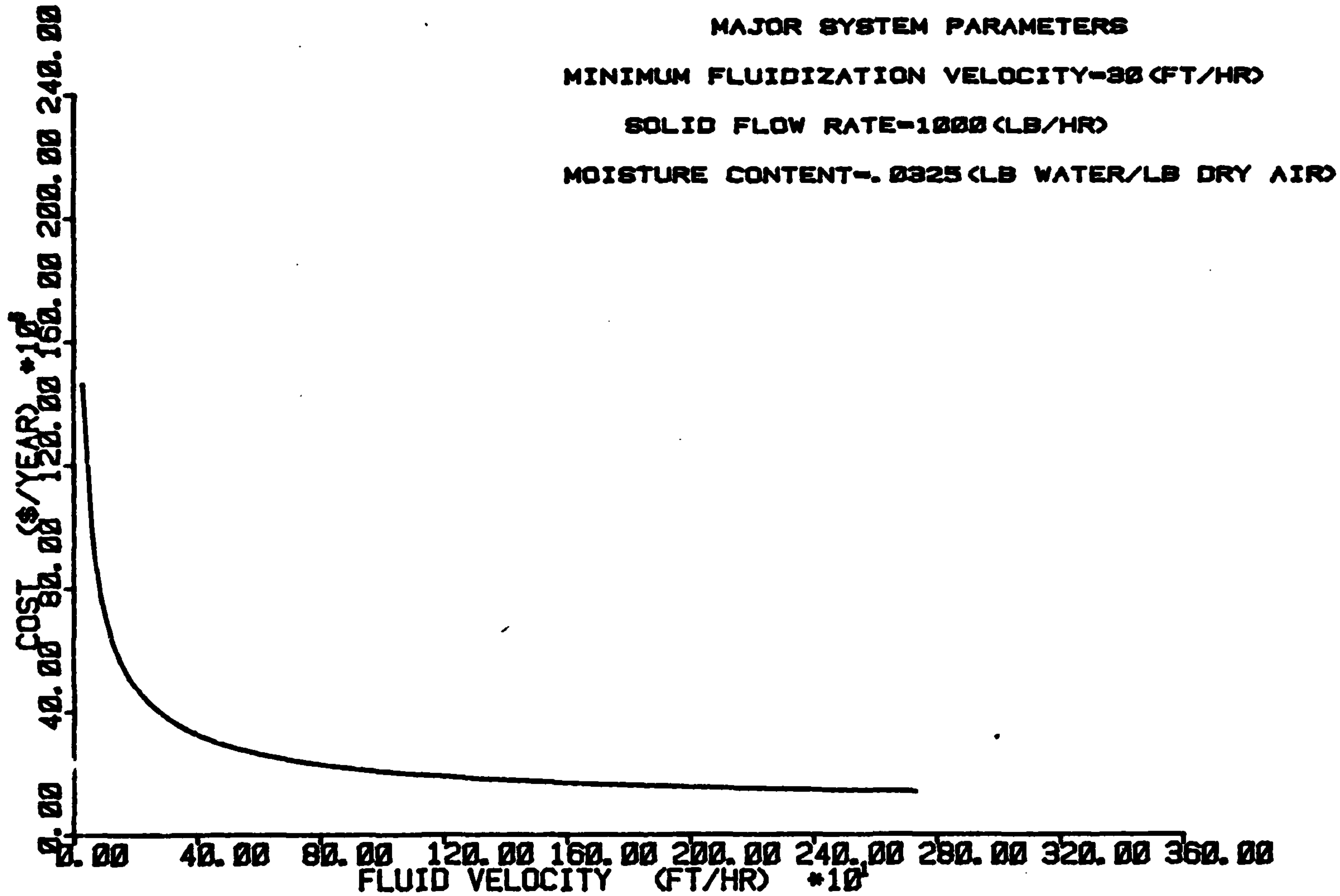

FIGURE (8). CAPITAL AND POWER COST

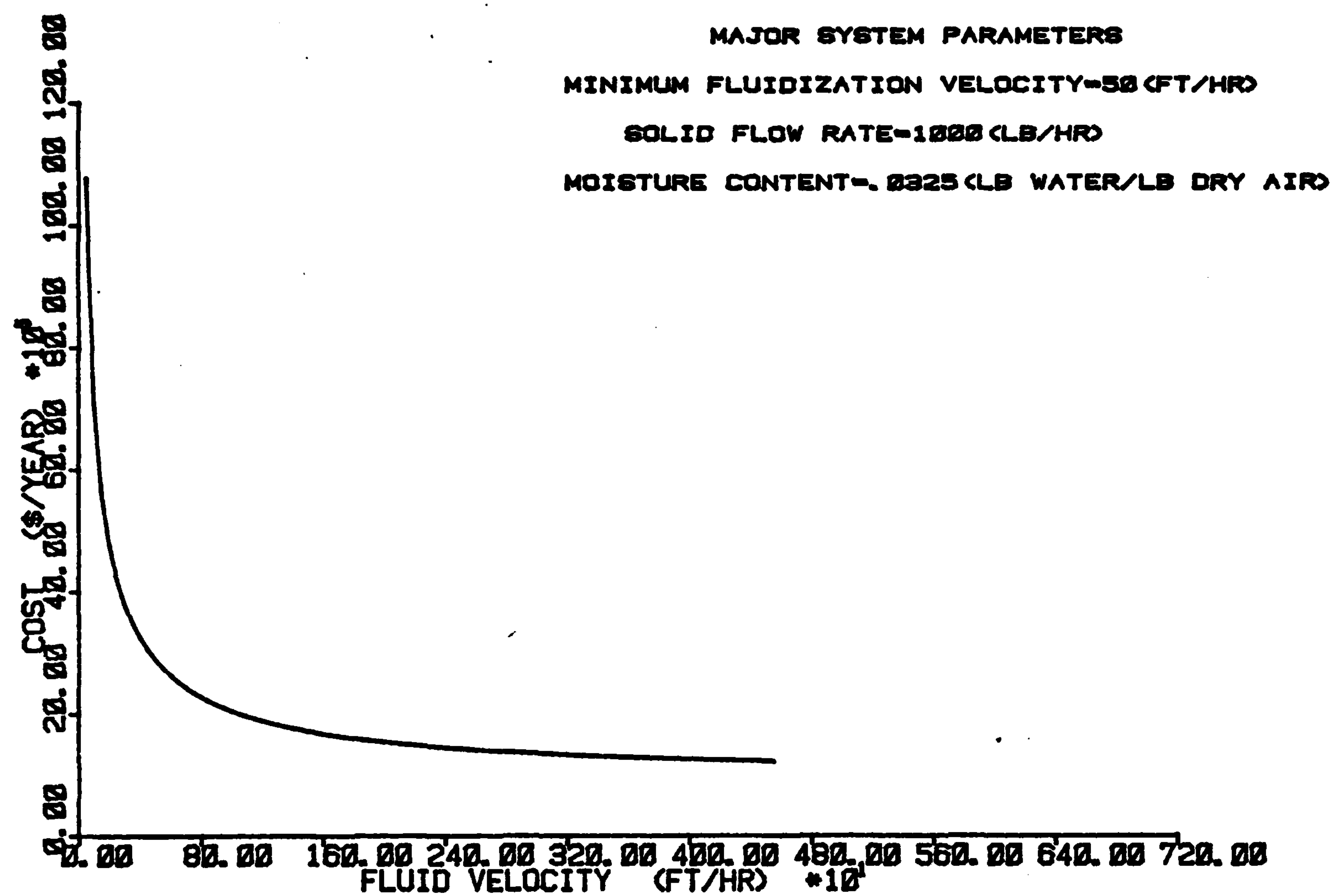

FIGURE (9). CAPITAL AND POWER COST

2. Solid Flow Rate

A change in the solid flow rate will affect each component of the cast model. The reasons for this are:

1. Dryer costs will increase with the solid flow rate increase due to the change in its design. The dryer needed for higher solid flow rate will have a greater volume and a greater diameter, consequently this greater diameter increases the peripheral area which is the determining factor in dryer capital costs.
2. Compressor costs will also increase since a higher pressure drop will be needed to fluidize the solid material. This requires a higher horsepower which leads to a higher capital cost.
3. The heater cost will increase with the solid flow rate increase due to the increase in the total moisture of the solid material. This requires more heating load which increases the peripheral area of the heater and consequently its capital cost.
4. Power consumption increases with a greater solid flow rate due to the increase in pressure drop needed for the fluidization process. This increases the electrical power consumption. Also the increased heating load required to remove the increased amount of solid moisture content increase the steam consumption.

Table 5 shows the optimal velocity change and the minimum cost change due to the changes in the solid flow rate. If the system were designed to operate at twice the solid flow rate, a decrease in optimal fluidization velocity by about 14 per cent which causes an increase in the minimum cost of 1 million or 90.8 per cent. At high flow rates of 7000 (lb/hr) the optimal velocity decreases to about one-half the optimum velocity of the base case (12,202 (ft/hr) while the cost increases by about 5.7 million or 513 per cent.

Table 5

Capital and Power Cost at Different

Solid Flow Rates

Case	Solid Flow Rate (lb/hr)	Optimal Velocity Range (ft/hr)		Difference from Base Optimal Velocity	Minimum Cost Range $\times 10^3$ ($/yr)		Cost Increase ($/yr)	Percent Cost Increase
		From	To		From	To		
1	Base (1000)	12,202	12,345	-	1,107.77	1,107.78	-	-
2	1000	10,483	10,340	-1719	2,113.21	2,113.24	1,005.44	90.8
3	3000	9,624	9,481	-2578	3,083.27	3,083.29	1975.5	178.3
4.	4000	9,051	8,908	-3151	4,031.01	4,031.02	2923.24	263.9
5.	5000	8,621	8,478	-3581	4,962.49	4,962.50	3854.72	348.0
6.	6000	8,192	8,335	-4010	5,881.18	5,881.28	4773.51	430.9
7.	7000	7,905	8,048	-4297	6,789.41	6,789.51	5681.64	512.9

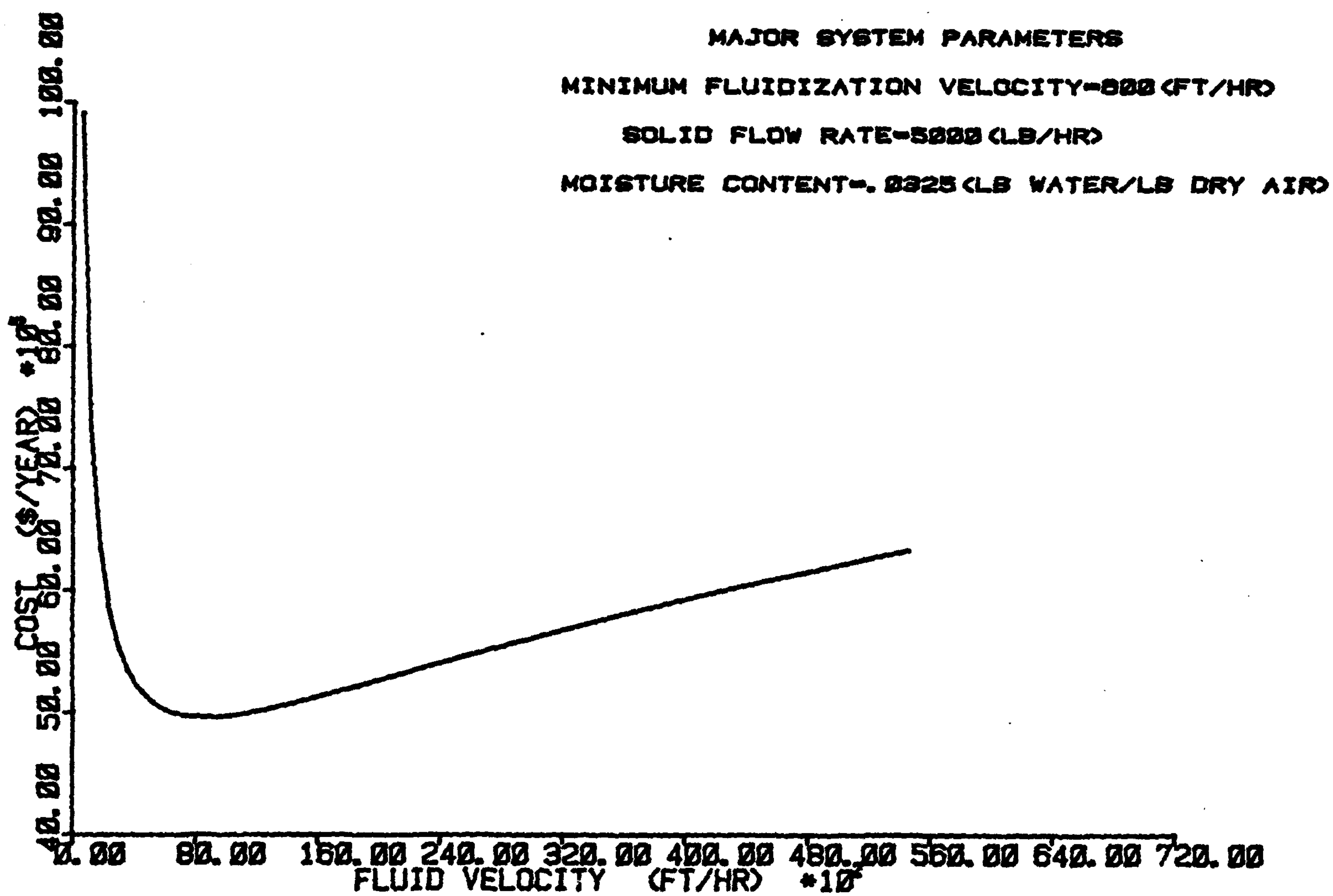

FIGURE (10). CAPITAL AND POWER COST

3. Other Auxiliary Factors

Other factors that have a slight influence on the cost model can be divided as major factors and minor factors. Major factors are the density of solid material and the moisture content of the outlet air while minor factors are the steam cost, the electricity cost, and the annual fixed-change factor. Effects of variation in these factors on model results are shown in Table 5. The essential effects are as follows.

1. Variation in solid density will require adjustments in the pressure drop and consequently in the compressor and its capital cost and the electrical power consumption of the process. For example, an increase in solid density to 150 (lb/cu.ft) results in a decrease in the optimum velocity range, a system design which includes a compressor of greater horsepower and an increase in system cost of about $300 thousands.
2. Variation in moisture content of outlet air from 0.0325 to 0.023 (lb.water/lb.dry air) has a great effect on the optimal fluidization velocity and a decrease in system cost of about $500 thousands. The optimal fluidization velocity reaches the terminal velocity in this case. Decrease the outlet moisture content means increasing in dry-bulb temperature of air as indicated in air-water

humidity charts (14). The effect of decreasing outlet moisture content of air is an increase in the optimal fluidization velocity and consequently an increase in the compressor cost. On the other hand, the dry-bulb temperature is increased which increases the logarithmic mean temperature and consequently the dryer peripheral area is decreased which decreases the dryer cost. This decrease in dryer cost is higher than the increase in compressor cost, so the net result of this case is a decrease in total cost.

3. Variation in the steam cost from 4.8 to 1.6 ($\$/10^6$ Btu) has no effect upon the optimal velocity range. However it increases or decreases slightly the capital and power cost.
4. Variation in the electricity cost from 0.48 to 0.16 ($/Kw. hr) has no effect upon the optimal velocity range. The capital and power cost will increase or decrease slightly more than was the case of variation in the steam cost.
5. Variation in the annual fixed change factor from 0.4 to 0.6 will have no impact on the optimal velocity range. Of course variation in this change will increase or decrease total cost.

Table 6

Capital and Power Cost at Different

Auxiliary Factors

Case	Factor to be Changed		Optimal Velocity Range (ft./hr.)		Minimum Cost Range x10^3 (\$/yr.)	
	Name	Value	From	To	From	To
1	Base Case		12,202	12,345	1,107.77	1,107.78
2	Solid density (lb./cu. ft.)	150	8,541	6,953	1,405.99	1,405.65
3	Moisture content of the outlet air (lb. water/ lb. dry air)	0.023	54,457	54,600	1,056.14	1,054.96
4	Steam Cost (\$/10^6 Btu.) Cs	1.6	12,202	12,345	1,105.81	1,105.82
5	Electricity Cost (\$/Kw.hr.) C_e	.16	12,202	12,345	1,105.07	1,105.07
6	Annual Fixed Change Factor R	0.6	12,202	12,345	1,326.53	1,326.53

B. Separate Cost Items

In this section the behavior of each of the component cost items is briefly reviewed.

1. Capital Cost

Capital cost represents 98.7 per cent of the total cost and it is obvious that the behavior of the total cost is mainly due to the behavior of capital cost. Table 7 and Figure 11 illustrate the behavior of capital cost. A comparison of Table 7 with Table 3 of the previous section shows that the optimal velocity range is the same, and the slight difference in the minimum cost range is due to the presence of power cost in Table 3.

Capital cost components are the dryer cost, the compressor cost, and the heat exchanger cost. Examination of the component capital cost shows the following:

1. As shown in Table 8 and Figure 12 the minimum cost dryer design is at the upper limit of operating velocity which is the terminal velocity.
2. Compressor capital cost increases as fluidized velocity increases as can be seen in Table 9 and Figure 13.
3. The heat exchanger capital cost is independent of velocity, has a fixed value at 961 ($/yr), and is very low compared with compressor or dryer capital costs representing about 0.1 per cent of total capital cost.

The optimal drying operation design results in a kind of compromise between the dryer and compressor costs corresponding to an optimal velocity intermediate between the minimum and terminal fluidization velocities of the system.

Table 7

Capital Cost

and Different Minimum Fluidization Velocities

Case	Minimum Fluidization Velocity (ft/hr.)	Optimal Velocity Range (ft/hr.)		Minimum Cost Range $\times 10^3$ (\$/yr.)	
		From	To	From	To
1	200	12,375	12,327	1,093.77	1,093.77
2	600	12,345	12,489	1,093.77	1,093.78
3	1000	12,459	12,220	1,093.77	1,093.78

Table 8

Dryer Cost

At Different Minimum Fluidization Velocities

Case	Minimum Fluidization Velocity (ft/hr.)	Optimal Velocity Range (ft/hr.)		Mimimum Cost Range $\times 10^3$ (\$/yr.)	
		From	To	From	To
1	200	18,152	18,200	295.42	294.94
2	600	54,457	54,600	151.48	151.23
3	1000	90,761	91,000	111.09	110.86

Table 9

Compressor Cost

at Different Minimum Fluidization Velocities

Case	Minimum Fluidization Velocity (ft/hr)	Optimal Velocity Range (ft/hr)		Minimum Cost Range x10^3 ($/yr)	
		From	To	From	To
1	200	248	200	201.71	187.84
2	600	743	600	290.18	270.41
3	1000	1239	1000	343.19	319.92

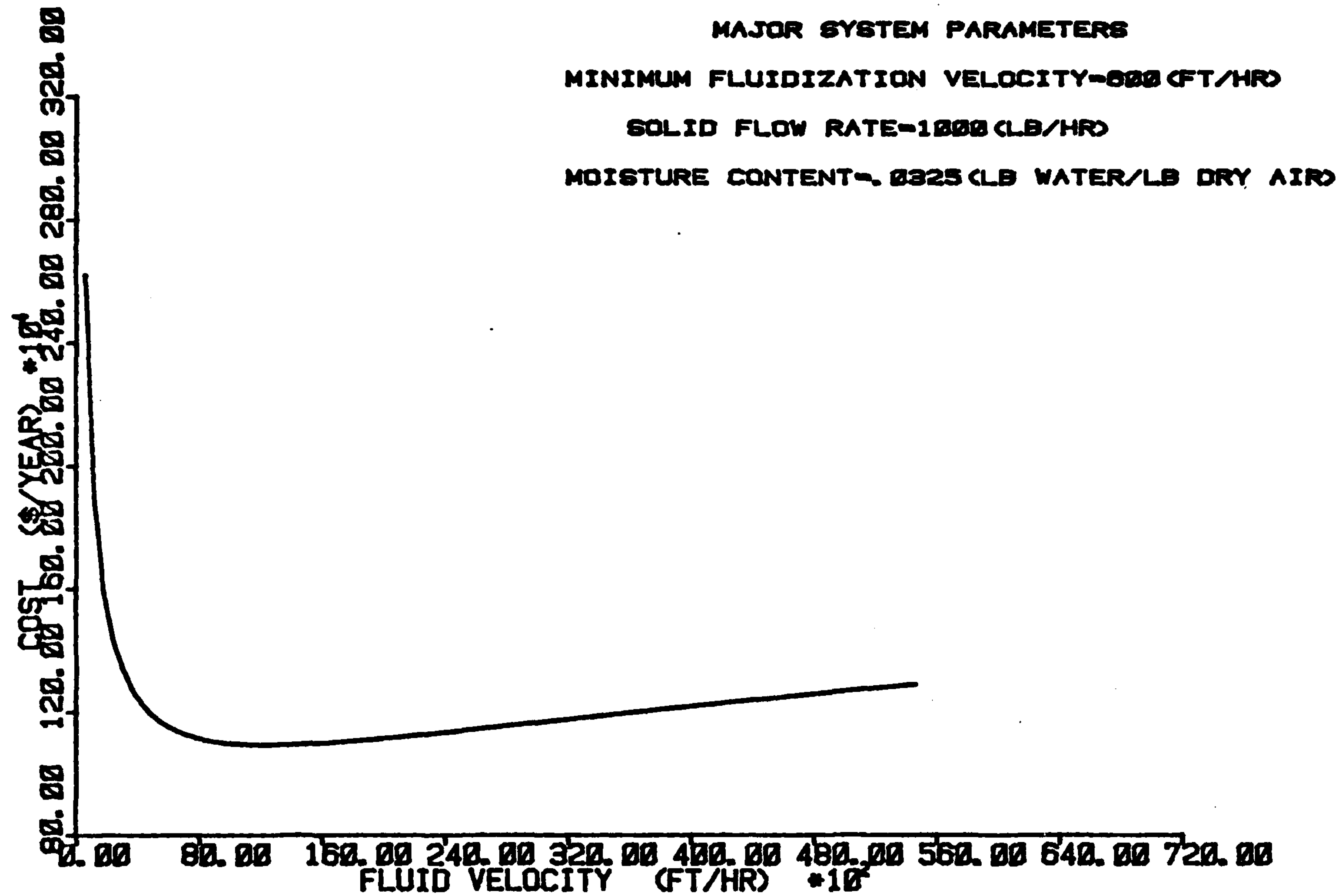

FIGURE (11). CAPITAL COST

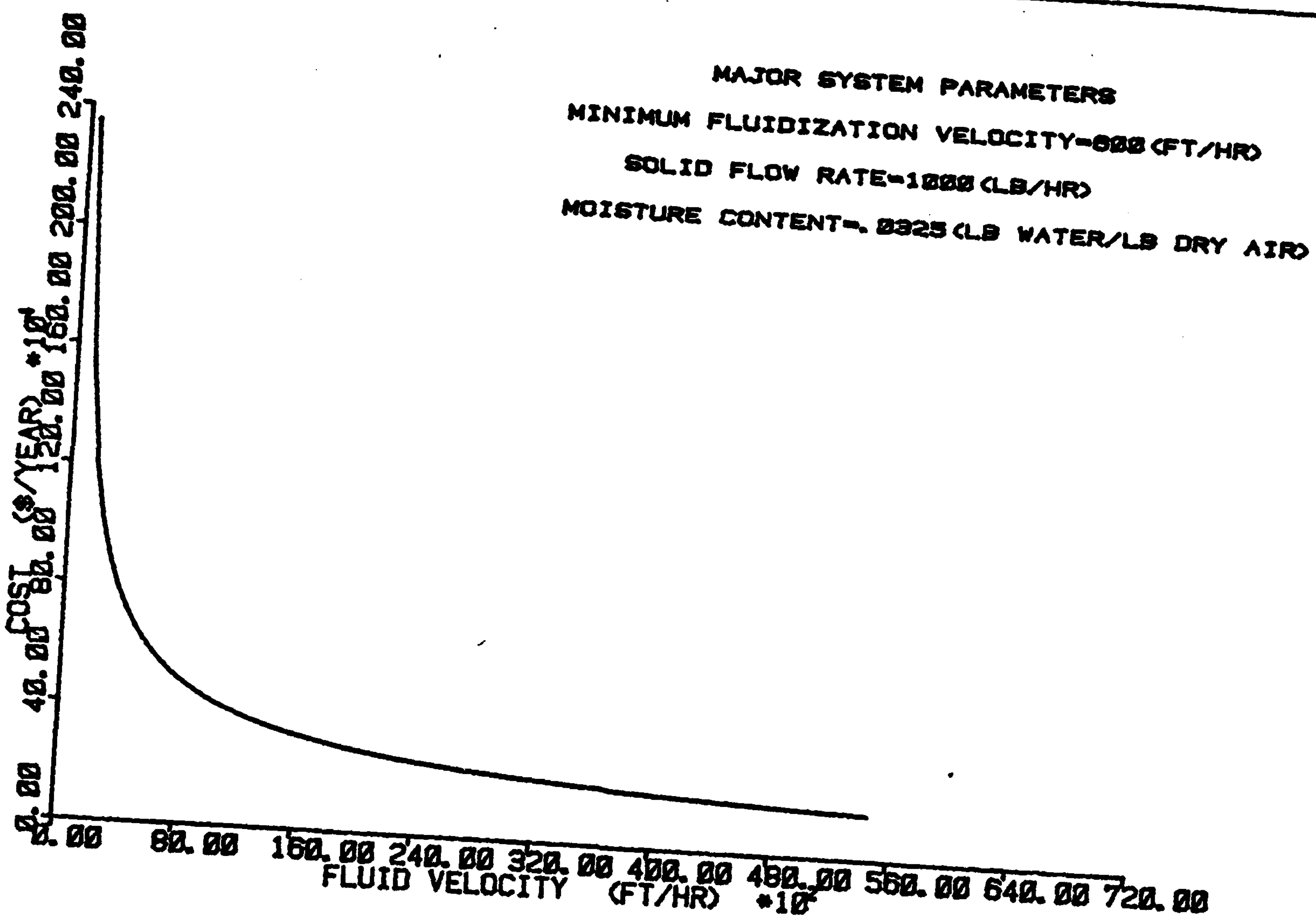

FIGURE (12). DRYER COST

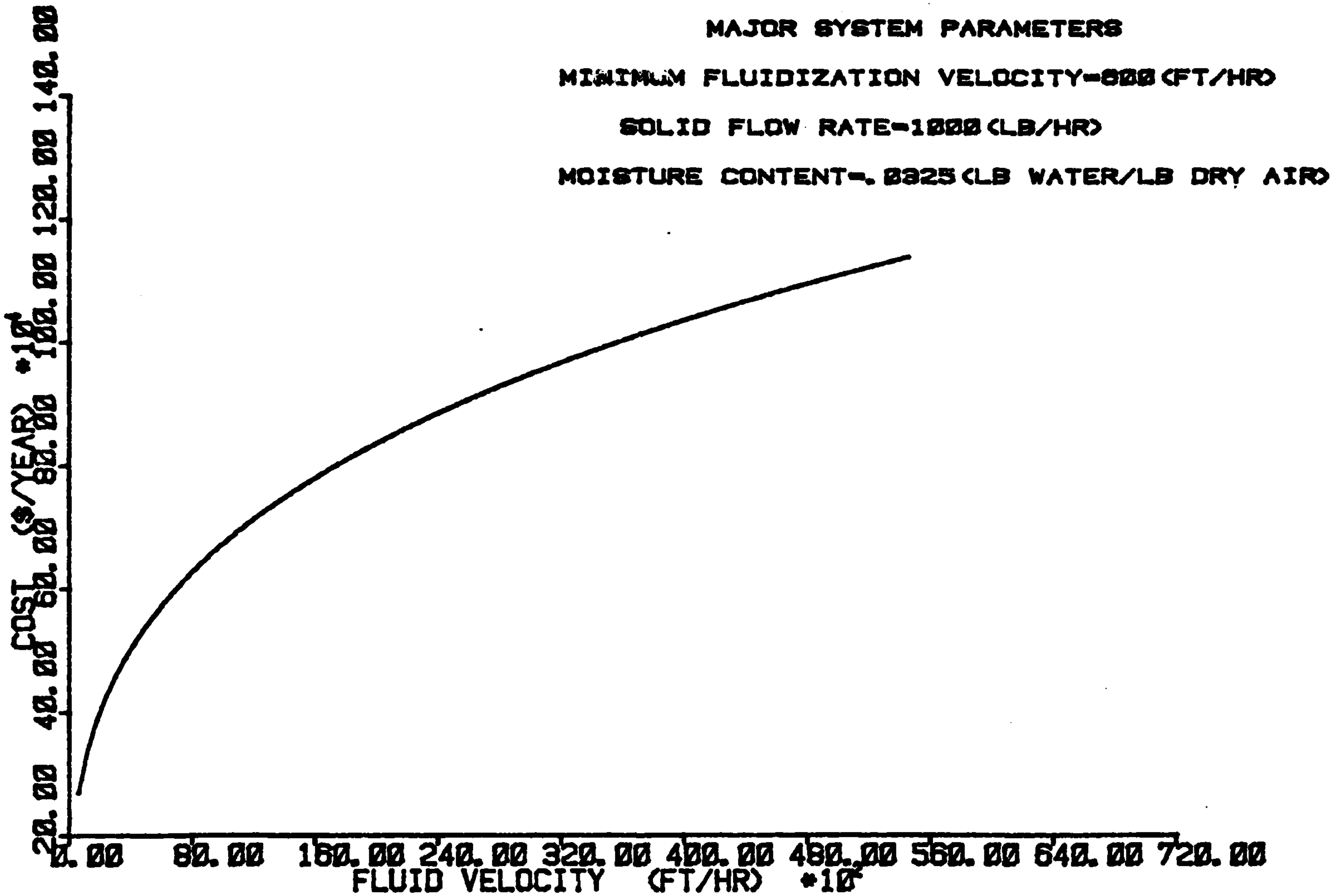

FIGURE (13). COMPRESSOR COST

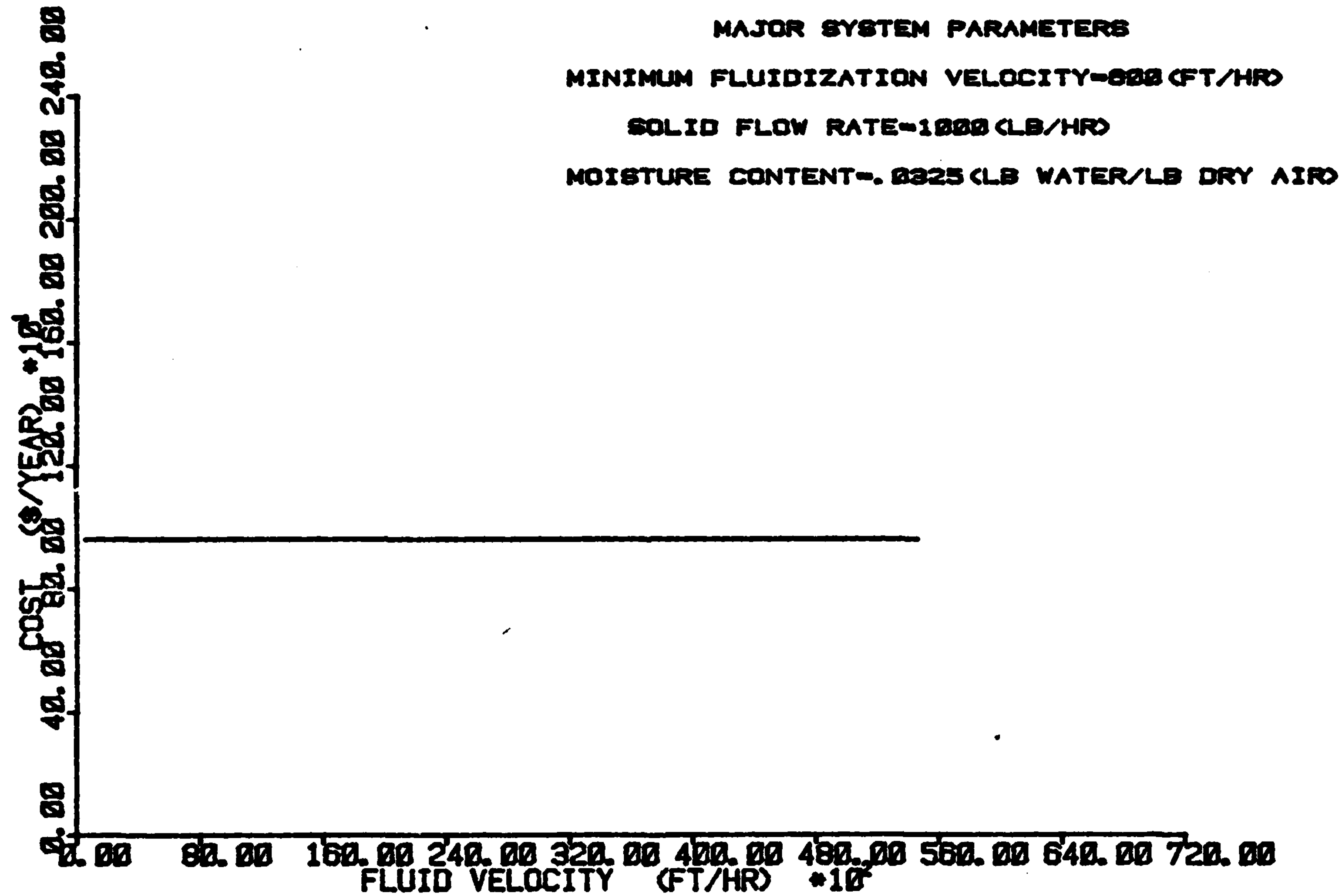

FIGURE (14). HEATER COST

2. Power Consumption Cost

The power consumption cost represents 1.3 per cent of the total cost. The components of power cost are the steam cost and the electricity cost. The behavior of power consumption cost and its components is shown in Figures 15, 16 and 17. As can be seen:

1. The power consumption cost increases with velocity
2. Steam cost is independent of velocity, and
3. The electrical consumption cost is responsible for the direct relationship of power cost to velocity.

The sensitivity of power cost to changing values of steam C_s and electricity C_e is shown in Table 10. As can be seen a doubling or halving of evergy price raises (reduces) power costs 30 (30) per cent. Optimal velocity and dryer design is little affected by small variation as noted above in Section A.

Table 10

Power Consumption Cost

at Different Steam and Electricity Prices

C_e	C_s	Optimal Velocity Range (ft/hr)		Minimum Cost Range $.10^3$ ($/yr)	
		From	To	From	To
.16	1.6	743	600	6.107	5.958
.16	2.4	743	600	8.067	7.918
.24	2.4	743	600	9.160	8.934
.24	1.6	743	600	7.201	6.978

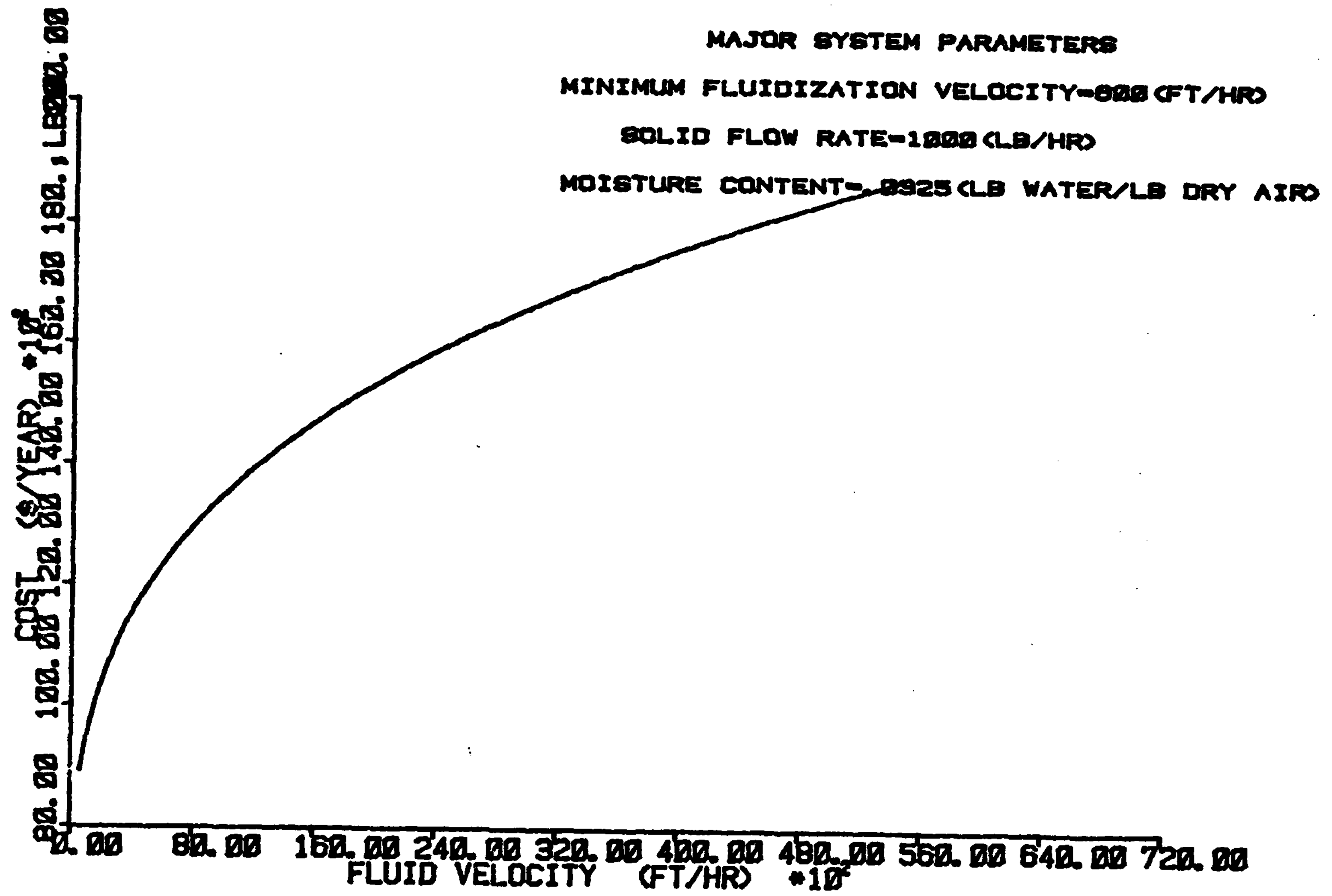

FIGURE (15). POWER CONSUMPTION COST

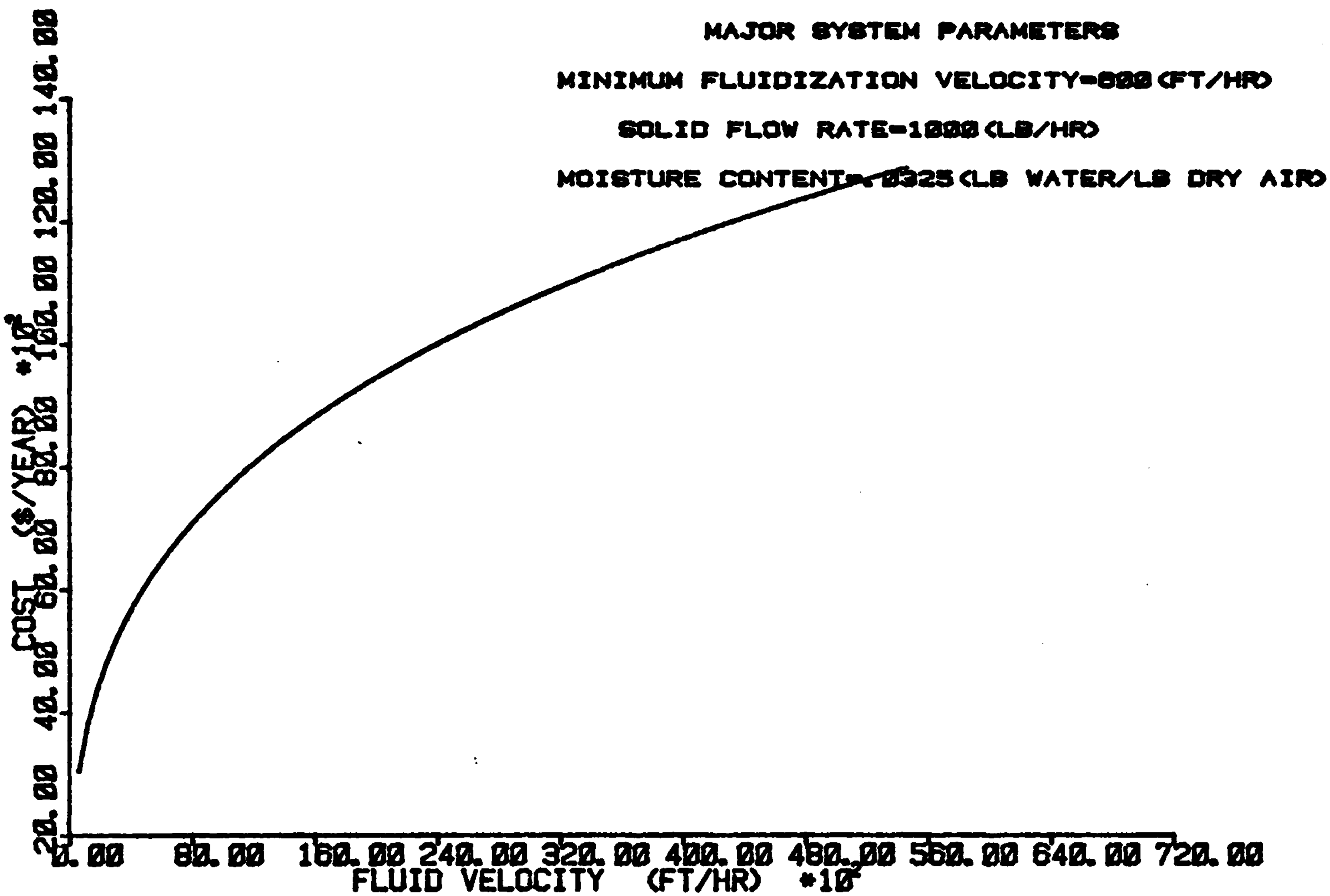

FIGURE (16). ELEC. CONSUMPTION COST

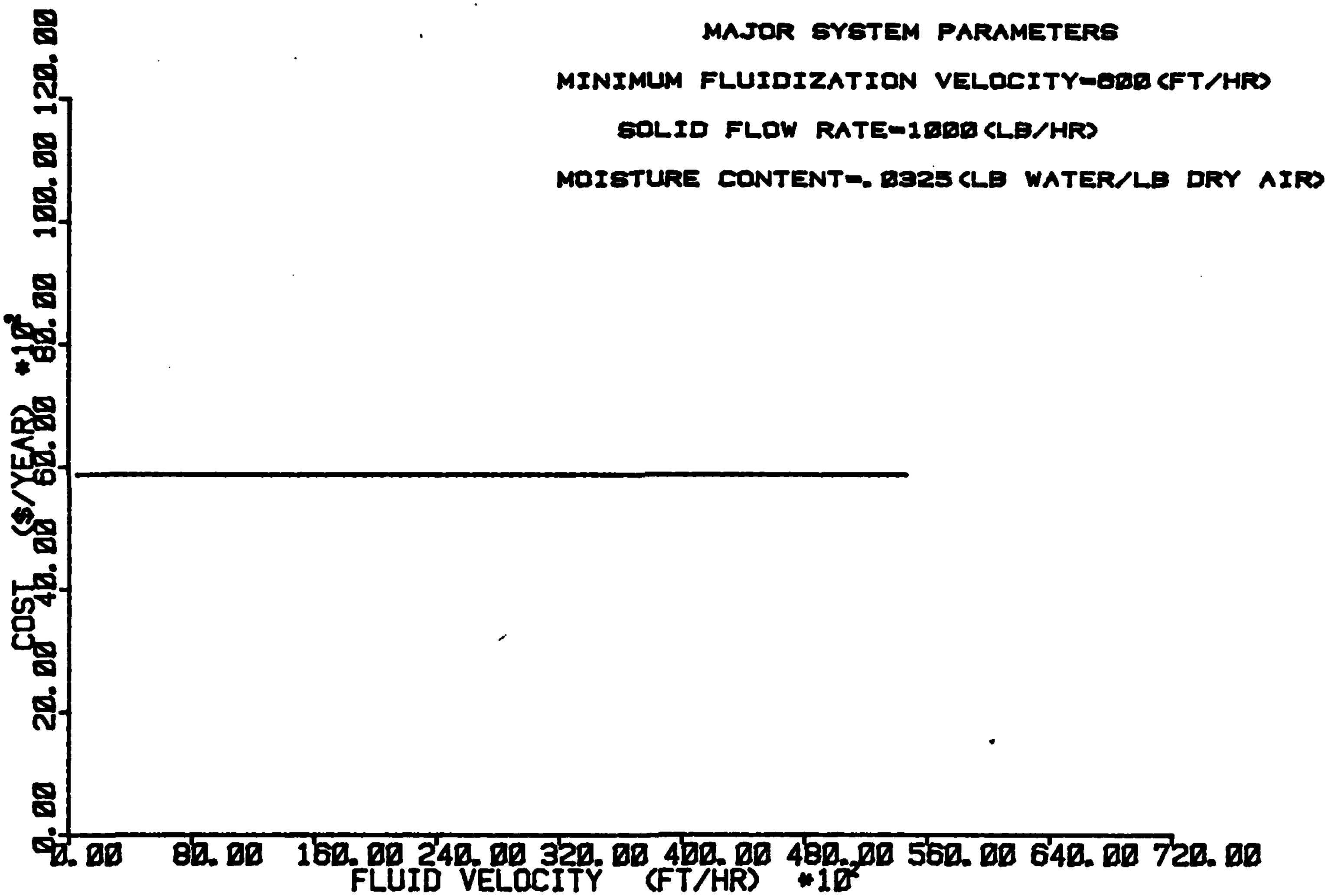

FIGURE (17). STEAM CONSUMPTION COST

C. Summary, Conclusions and Recommendations

The cost model developed and analyzed in this study can be employed to identify optimal fluidization velocities and their associated minimum cost equipment designs. The model results can be summarized as follows:

1. Capital costs are very high compared to power costs. The former represents 98.7 per cent of the total cost of an optimum dryer design configuration.
2. Dryer capital cost varies inversely with fluid velocity and represents 33 - 37 per cent of the capital cost of an optimal design configuration.
3. Compressor costs vary directly with fluid velocity and represent 63 - 67 per cent of the capital cost of an optimal design.
4. The optimal capital design requires a dryer, compressor, and heater configuration which has an optimal fluidized velocity which is intermediate to the minimum fluidization velocity (which gives lowest compressor cost) and the terminal velocity (which gives lowest dryer costs).
5. Heater requirements increases the capital costs of the optimum design at the indicated optimal velocity level.
6. The power cost is relatively unimportant representing 1.3 per cent of total design cost. Of this, steam requirements represents 42 per cent of the power

cost, while electricity requirements represents 58 per cent of total power cost. Only the latter varies with fluidization velocity, increasing as the pressure drop increases.

7. Equipment designs outside the optimal fluidization velocity range impose cost penalties. Cost penalties can be held to less than 4 per cent of system costs so long as design configurations correspond to fluid velocities no less than one half the optimal velocity nor more than double the optimal velocity. Very severe cost penalties are imposed on system designed at or close to minimum fluidization velocity levels.

A brief check in the validity of the design model is possible by checking two key aspects of the model against reported real world drying operations. First the real world examples of fluidized velocity drying operation report velocity levels which lie inside the optimal velocity range identified by the model (11), (8).

Second the height/diameter ratio reported to be most efficient in commercial practice lies between 4 and 10 (15). Solutions of the model for a solid flow rate from 7,800 to 10,000 (lb/hr) and fluidized velocity in the optimal range 5,000 to 8,000 (ft/hr) yields a ratio of height/diameter of 4 to 5. These limited information on operating systems is agreed with the results of our model.

The model developed proposed in this study represents a first step in the development of a mathematical model for use in the design and study of fluidized-bed drying operations. Model modification to include the height-diameter ratio as a constraint in the design process would be a useful extension of this model. Also the mechanisms of fluidized-bed bubble formation and/or channeling and the mechanism effect on heat transfer might be usefully considered in future development of the model.

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APPENDIX A

Fibonacci Algorithm

This algorithm finds the minimum of a single variable, nonlinear function subject to constraints.

Minimize $F(X)$

subject to $a_1 \leq x \leq b_1$

The upper and lower bounds, b_1 and a_1 are constants.

The procedure is an interval elimination search method. Thus, starting with the original boundaries on the independent variable, the interval in which the optimum value of the function occurs is reduced to some final value, the magnitude of which depends on the desired accuracy. The location of points for functions is based on the use of positive integers known as the Fibonacci numbers. No derivatives are required. A specification of the desired accuracy will determine the number of function evaluations. A unimodal function is assumed. The algorithm proceeds as follows:

1. Designate the original search interval as L_1 with boundaries a_1 and b_1.
2. Predetermine the desired accuracy α and thus the number, N, of required Fibonacci numbers.

$$\alpha = \frac{1}{F_N}$$

$$F_o = F_1 = 1$$

$$F_n = F_{n-1} + F_{n-2}, \; n \geq 2$$

where F_n is called a Fibonacci number

3. Place the first two points, X_1 and X_2 ($X_1 < X_2$) within L_1 at a distance l_1 from each boundary.

$$l_1 = \frac{F_{N-2}}{F_N} L_1$$

$$X_1 = a_1 + l_1$$

$$x_2 = b_1 - l_1$$

4. Evaluate the objective function at X_1 and X_2. Designate the functions as $F(X_1)$ and $F(X_2)$. Narrow the search interval as follows:

$$a_1 \leq X^* \leq X_2 \qquad \text{for } F(X_1) < F(X_2)$$

$$X_1 < X^* \leq b_1 \qquad \text{for } F(X_1) > F(X_2)$$

Where X* is the location of the optimum. The new search interval is given by

$$L_2 = \frac{F_{N-1}}{F_N} \cdot L_1 = L_1 - l_1$$

with boundaries a_2 and b_2.

5. Place the third point in the new L_2 subinterval, symmetric about the remaining point,

$$l_2 = \frac{F_{N-3}}{F_{N-1}} L_2$$

$$X_3 = a_2 + l_2 \text{ or } b_2 - l_2$$

6. Evaluate the objective function $F(X_3)$, compare with the function for the point remaining in the interval and reduce the interval to

$$L_2 = \frac{F_{N-2}}{F_N} L_1 = L_2 - l_1$$

7. The process is continued per the preceeding rules for N evaluations.

$$l_k = \frac{F_{N-(k+1)}}{F_{N-(k-1)}} L_k$$

$X_{L+1} = a_k + l_k$ or $b_k - l_k$ (symmetric about mid point)

$$L_k = \frac{F_{N-1\ (k-1)}}{F_N} L_1 = L_{k-1} - l_{k-1}$$

After N-1 evaluations and discarding the appropriate interval at each step, the remaining point will be precisely in the center of the remaining interval. Thus point N is also at the midpoint and is replaced by a point perturbed some small distance ε to one side or the other of the midpoint. The objective function is then evaluated and the final interval where the optimum is located is thus determined.

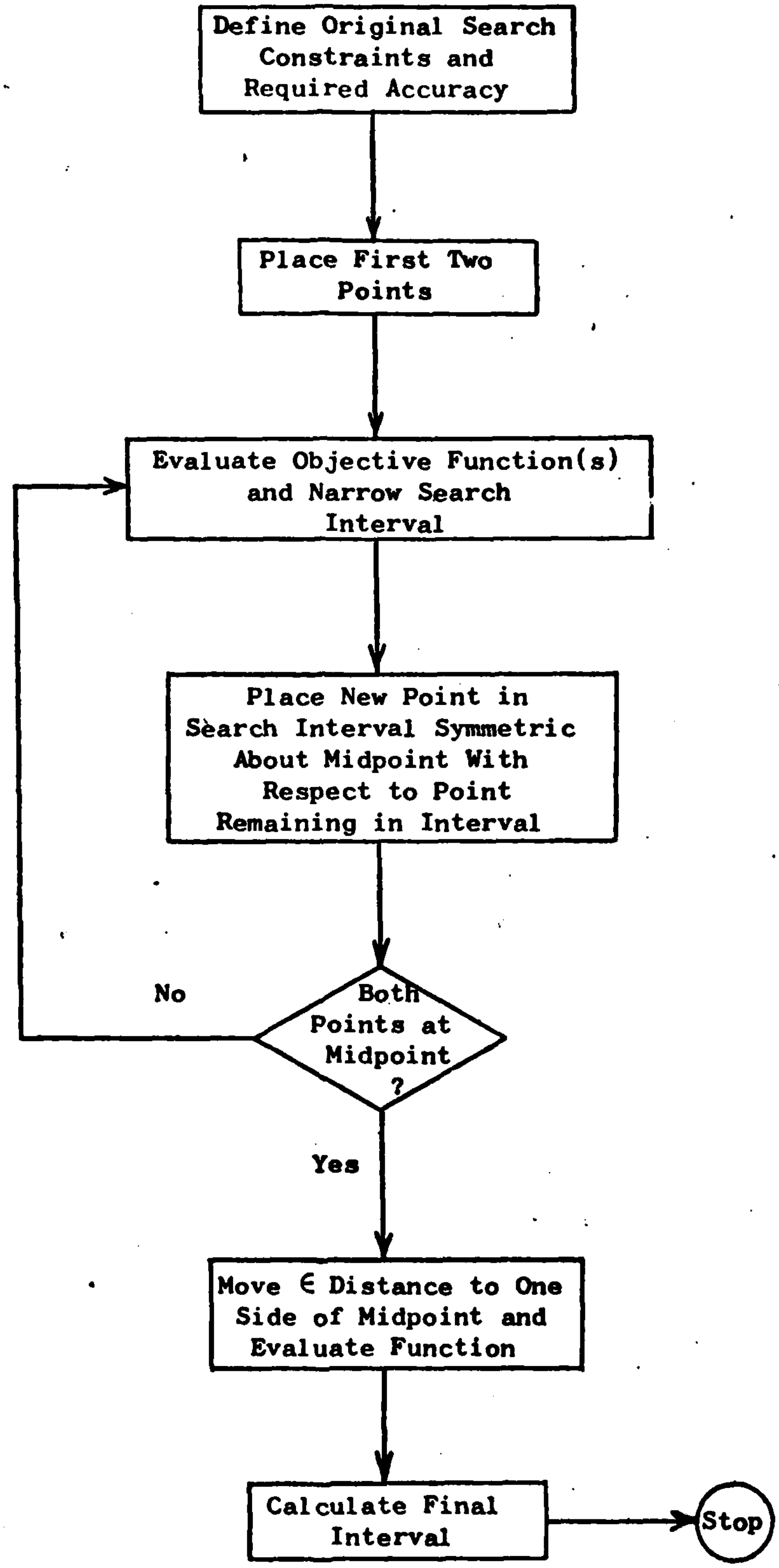

Figure 18 Fibonacci algorithm logic diagram

APPENDIX B

```
***************************************************************
*  THIS A FORTRAN CODE OF FIBONACCI SEARCH ALGORITM USED IN   *
*  FINDING THE OPTIMUM FLUIDIZATION VELOCITY OF FLUIDIZED-BED *
*  DRYING OPERATION.                                          *
***************************************************************
          PROGRAM FIBONA
          DIMENSION FIB(50)
          ALPHA=.01
          A=600
          A1=A
345       B=91*A
          DEL=B-A
          WRITE(*,001)
001       FORMAT(//////)
567       FORMAT(15X,'* MINIMUM FLUIDIZATION VELOCITY=',F5.0,'(FT/HR)*')
223       FORMAT(22X,'* HO=.0325,TDO=95.,W=1000 *')
789       FORMAT(22X,'* CAPITAL AND POWER COST *')
          WRITE(*,567)A
          WRITE(*,223)
          WRITE(*,789)
C
C         DEFINE THE FIRST THREE FIBONACCI NUMBERS
C
          FIBO=1.0
          FIB(1)=1.0
          FIB(2)=2.0
C
C         CALCULATE THE REMAINING FIBONACCI NUMBERS
C
5         BB=1.0/ALPHA
          IF(BB-2.0)10,10,11
10        GO TO 14
11        CONTINUE
          JJ=2
12        JJ=JJ+1
          FIB(JJ)=FIB(JJ-1)+FIB(JJ-2)
          CC=FIB(JJ)
          IF(CC-BB)13,15,15
13        GO TO 12
14        WRITE(*,002)
002       FORMAT(///,10X,'ACCURACY SPECIFIED IN FUNC NOT SUFFICIENT',
         1//,10X,'PROGRAM RESET ALPHA,ALPHA=.01')
          ALPHA=.01
```

```
      GO TO 5
C
C     FIRST STEP IN THE TABLEAU
C
15    I=0
      KK=JJ-2
      IK=JJ-2
      BL=B-A
      ALL=FIB(IK)*BL/FIB(JJ)
      W=A+ALL
      V=B-ALL
      CALL FUNC(W,T)
      CALL FUNC(V,U)
      JK=1
      WRITE(*,555)
      WRITE(*,777)
      WRITE(*,003)
003   FORMAT( '**','K',' *',4X,'LK',4X,'*',4X,'AK',4X,'*',
     +4X,'BK',4X,'*',3X,'LLK',4X,'*',' VELOCITY *',' TOTAL COST *' )
777   FORMAT(1X,'*',2X,'*',10X,'*',10X,'*',10X,'*',10X,'*',
     + 10X,'*',12X,'*')
      WRITE(*,777)
      WRITE(*,555)
      WRITE(*,777)
      WRITE(*,004)JK,BL,A,B,ALL,W,T
      WRITE(*,777)
      WRITE(*,555)
      WRITE(*,777)
004   FORMAT( '**',I2,'*',E10.4,'*',E10.4,'*',E10.4,'*',E10.4,
     1 '*',E10.4,'*',E12.6,'*' )
006    FORMAT(51X,'*',E10.4,'*',E10.4)
C
C     SUCCEEDING STEPS IN THE TABLEAU
C
      IK=IK-1
      JJ=JJ-1
      DO 70 I=1,KK
      IF(U-T)20,20,22
20    A=A+ALL
      BL=B-A
      W=V
      CALL FUNC(W,T)
      ALL=FIB(IK)*BL/FIB(JJ)
      V=B-ALL
      CALL FUNC(V,U)
      II=I+1
      IK=IK-1
      JJ=JJ-1
      IF(IK-1)28,29,29
28    IK=1
```

```
29    CONTINUE
      WRITE(*,004)II,BL,A,B,ALL,W,T
      WRITE(*,777)
      WRITE(*,555)
      WRITE(*,777)
      GO TO 70
22    B=B-ALL
      BL=B-A
      V=W
      CALL FUNC(V,U)
      ALL=FIB(IK)*BL/FIB(JJ)
      W=A+ALL
      CALL FUNC(W,T)
      II=I+1
      IK=IK-1
      JJ=JJ-1
      IF(IK-1)30,31,31
30    IK=1
31    CONTINUE
      WRITE(*,004)II,BL,A,B,ALL,V,U
      WRITE(*,555)
555    FORMAT(1X,72('*'))
      GO TO 70
70    CONTINUE
C
C     CALCULATION OF THE FINAL RANGE OF THE DEPENDENT VARIABLE
C
      EPS=.001*W
      DL=EPS+W
      CALL FUNC(DL,YL)
      IF(YL-T)80,80,81
80    CALL FUNC(B,BF)
      WRITE(*,333)
333   FORMAT(////,2X)
      WRITE(*,567)A1
      WRITE(*,223)
      WRITE(*,789)
      WRITE(*,666)
      WRITE(*,444)
444   FORMAT(1X,'*',24X,'*',15X,'*',15X,'*')
      WRITE(*,007)W,B
      WRITE(*,444)
007   FORMAT('** THE OPITMAL VELOCITY   *',2X,E13.7,'*',2X,E13.7,'*'
      WRITE(*,666)
      WRITE(*,444)
666   FORMAT(1X,58('*'))
      WRITE(*,008)T,BF
      WRITE(*,444)
008   FORMAT('** CAPITAL AND POWER COST *',2X,E13.7,'*',2X,E13.7,'*'
      WRITE(*,666)
      WRITE(*,333)
```

```
      GO TO 87
81    CALL FUNC(A,AF)
      WRITE(*,007)W,A
009   FORMAT(///,'  THE FINAL FEASIBLE REGION',2X,'X=',
     1 E13.7,2P,'X=',E13.7)
      WRITE(*,008)T,AF
      WRITE(*,333)
017   FORMAT(/'  WITH FUNCTION VALUES',7X,'Y=',E13.7,2X,'Y=',
     + E10.4)
87    ACC=(W-A)/(DEL)
999   CONTINUE
123    END
C
      SUBROUTINE FUNC(X,Y)
      W=1000.0
      XI=.111
      XO=.0101
      TDI=200
      TDO=95
      TW=92
      HI=.0008
      ROH=.074
      DELM=((TDI-TW)-(TDO-TW))/(ALOG(TDI-TW)/(TDO-TW))
      EP=.4
      ROS=150
      P2=14.7+.4*((ROH*X)**.34)*(1-EP)*(ROS-ROH)/DELM
      HO=.0325
      UH=5
      TGO=200
      TGI=70
      TS=250
      DELH=((TS-TGI)-(TS-TGO))/(ALOG(TS-TGI)/(TS-TGO))
      SPCR=.29
      R=.5
      CE=0.24
      CS=2.4/.1E+07
      DRCT=    R*(585*((200*W*(XI-XO))/(ROH*X)**.76*DELM)**.8)
      HTCT=R*(346*(W*(XI-XO)*(.24*(TGO-TGI)+
     + .5*HI*(TGO-TGI))/((HO-HI)*UH*DELH))**.62)
      COCT=R*5873.63*(W*(XI-XO)*SPCR*((P2/14.7)**SPCR-1)/((ROH*(
     + HO-HI)))**.8)
      STCT=8000*CS*(W*(XI-XO)*(.24*(TGO-TGI)+
     + .5*(TGO-TGI))/(HO-HI))
      POCT=41018.88*CE*W*(XI-XO)*SPCR*((P2/14.7)**SPCR-1)/
     + ROH*(HO-HI)
      CAPCT=DRCT+HTCT+    STCT+POCT+COCT
      Y=CAPCT
      RETURN
      END
*EOR
```

VITA

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20. ABSTRACT (*Continue on reverse side if necessary and identify by block number*)

This research investigates the economic characteristics of a fluidized-bed drying process. It focuses on the way in which fluid velocity impacts the capital and power consumption costs of such a system.

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The engineering and economic relationships employed in the model are developed and a sensitivity of the optimal design of a fluidized-bed drying process based on capital and operating cost is explored through variation in the technical parameters of the model.

END