FUNDAMENTAL STUDIES IN FLASH DRYING

by

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INTRODUCTION

The drying of moisture from materials is one of the oldest of operations. It was initially used in connection with food preservation, but today it finds application in almost every industrial process.

Several types of driers have been developed and are generally classified as rotary, tunnel, spray, or drum types. In common driers using air as the drying medium, the latent heat of vaporization is supplied to the wet material by convection from the heated air. The drum type of drier, however, supplies the heat for vaporization of water by conduction from a heated metal drum. In such driers, instead of the material receiving heat from the air, the air receives heat as the water is vaporized. Each type of drier has severe limitations and is confined to a narrow and specific field of application. Commercial drying equipment, therefore, is extremely varied in design because materials that are dried vary widely in both physical and chemical properties.

Flash drying is of comparatively recent origin. By its very nature flash drying creates conditions that produce rapid and efficient removal of moisture. This type of drying is suitable for all materials that can be dispersed in a gas stream and conveyed. Temperature sensitive materials can often be dried at high temperatures due to the short drying time and the fact that the particles are not allowed to exceed the wet-bulb temperature of the entering gas stream.

Despite the many good features of flash drying, there is

no published information on the fundamental theory. The purpose of this investigation is to relate the independent variables, so far as possible, to the rates of mass and heat transfer during flash drying. It is therefore intended to present the experimental data in such a way as to facilitate its use in the design of flash driers.

THEORY OF FLASH DRYING

The development of flash drying is of comparatively recent origin, dating back only to the early 1930's (5). Some of the basic concepts are quite simple and easily understood. By its very nature, flash drying creates conditions that produce rapid and efficient removal of moisture. The drier is essentially composed of a device for dispersing a wet solid into hot gases, a duct through which these gases convey the dispersed particles, and a collection system for removing the dry product from the air stream.

As the dispersed wet solid is conveyed through the drying section by the hot gas it is continually losing moisture. This loss of moisture is so great as compared to the slight change in volume that the density of the particle is greatly reduced. A decrease in the density of the particle also decreases the velocity or gas rate required to support the particle in the gas stream. The lower gas rate is achieved in part by increasing the diameter of the duct near the top of the drying section. As more moisture is lost by the particle it is floated gradually higher and higher in the column until the density has decreased

to such a value that it passes out of the drying column and into the separator. Because of the action of the particle this type of drying is commonly referred to as "floating bed flash drying". In most instances, because information is lacking for making the delicate balance of variables required for floating bed flash driers, a drying column of a constant diameter is commonly used. The drying time is then varied by altering the length of the drying section. This type of system, referred to as a "pneumatic conveying drier", has a constant throughput and does not have the classification feature displayed by the floating bed type of drier.

Lorenzi (5) lists four important factors which govern the extent, as well as the rapidity, of evaporation. They are:

1. Moisture dispersion.
2. Temperature differential.
3. Agitation.
4. Particle size.

The ideal situation would be to have all moisture concentrated close to the surface of the particle so as to facilitate the rapid drying. This case, however, is unusual as the condition usually exists where the moisture is dispersed throughout the volume of the entire particle. It is the interior moisture that frequently offers the major obstacle to rapid rates of drying not only in flash drying but also by the more conventional methods.

The drying of most materials takes place in two stages; first, a constant-rate period; and second, the falling-rate period. During the constant-rate period, liquid moisture is

transferred to the surface of the solid by capillary forces at a rate equal to that of the evaporation from the surface. During this period the surface of the solid is completely wetted, and drying is analagous to the evaporation from a free water surface. The rate of drying is determined by the rate of heat transfer to and the rate of diffusion of the water vapor from the particle surface through the gas film at the surface of the solid. A constant rate of evaporation on the surface of the solid tends to maintain the surface at a constant temperature, which in the absence of other heat effects, is the wet-bulb temperature of the air stream. When heat arrives at the surface of evaporation by radiation, in addition to conduction, this surface temperature will lie somewhere between the air temperature and the wet-bulb temperature and in turn will produce a higher rate of evaporation. Radiation is usually secondary to conduction, although in some cases it is a primary mechanism, as in infrared drying (6) or when drying at extremely high temperatures in gases having an absorption band in the infrared region.

As the moisture content of the particle diminishes, a point is reached at which the rate of movement of the moisture to the surface of the particle by capillarity is no longer equal to the rate of evaporation at the surface. This point is known as the "critical moisture content" of the solid. From this point on, the plane of vaporization gradually moves from the surface to the interior of the particle. The distance which this plane of vaporization moves in a given length of time depends upon several factors, such as the rate of liquid moisture diffusion through the

particle, the rate of water vapor diffusion out of the particle, and the rate of heat transfer into the particle. As the plane of vaporization begins to move into the particle the rate of evaporation diminishes and the particle tends to assume the temperature of the gas stream.

If the critical moisture content is less than the required final moisture content, the constant-rate period will constitute the entire drying process. Materials of low critical moisture content are particularly suited to flash drying since flash drying depends almost entirely on drying during the constant-rate period.

High gas temperatures are desirable for rapid drying as the rate of heat-transfer from the gas to the moisture in the particle of material being dried is proportional to the difference between the initial temperature and saturation temperature of the drying medium. Use of high temperatures is made possible by the fact that the product is not allowed to exceed the wet-bulb temperature. It is for this reason, provided the material is properly dispersed into the hot gas stream, that flash drying is particularly well adapted to the difficult task of drying organic materials whose nature may be such as to be subject to heat damage which would result in the destruction of valuable ingredients, or properties, and in the production of disagreeable odors (4).

The behavior of a flash drier is based upon certain psychrometric laws. For example, assume that an adiabatic drying operation will start with an entering gas stream temperature of

1000° F. dry-bulb, and 0.02 lbs. water per lb. bone dry air or a wet-bulb temperature of 150° F. The drying proceeds at the wet-bulb temperature of 150° F. to the point where the exhaust gas dry-bulb temperature is 250° F. The moisture content of the gas has now increased to 0.225 lb. water per lb. bone dry air, the gain representing the moisture picked up from the wet solids. The drop in the dry-bulb temperature from 1000° F. to 250° F. represents the exchange of sensible heat of the gases to the latent heat of vaporization for the moisture increase in the gas stream. Provided that the critical moisture content is not reached during this period of the heat interchange process, the solids being dried remain at the wet-bulb temperature of 150° F. Then as Lorenzi (5) states, the two principles upon which this type of operation is based are:

> 1. A body of heated gas will retain the same wet-bulb temperature while being partially or fully saturated with moisture, provided the total heat in the system remains constant.
>
> 2. A particle suspended in a body of hot gas assumes the wet-bulb temperature of the gas, provided the moisture content remains above the critical moisture content of the particle.

In actual practice, however, flash drying does not strictly adhere to these laws. Frequently the product temperature is above the wet-bulb temperature of the gas stream.

When a particle gives up moisture to a stagnant hot gas,

Adiabatic - a type of process in which there is no heat absorbed or given off by the system, hence the total change in heat content remains zero.

an envelope of vapor is formed around it. This envelope tends to prevent further evaporation by setting up a momentary equilibrium condition around the particle. Agitation, such as that which results in a flash drier by the rapid passage of the hot gas past the particles, overcomes this momentary condition by continuously reducing the thickness of the vapor film. Since in the drying process the rate of evaporation is determined by the rate of mass-and heat-transfer through the gas film, the reduction of the thickness of this film is of utmost importance. The drying rate is an inverse function of the vapor film thickness.

Particles of small diameter dry faster than particles of large diameter because of the greater available area for transfer. Furthermore, because the critical moisture content of any given material is lower for small particles than for large particles, smaller particles may be dried to a lower moisture value without heat damage. Thus, Gordon (4) states that operations which require simultaneous grinding and drying are particularly suited to flash drying.

According to Perry (6), there is no published information on the fundamental theory of the pneumatic conveying or floating bed flash drying. He states, however, that if the particle size is known and is such that the heat-transfer coefficient is independent of the conveying velocity, the drying time for each particle in a pneumatic conveyer type drier may be estimated by use of the following equation:

$$\theta = \frac{\rho_s (D_p^2)\ \lambda\ (W_o - W_1)}{12\, k_f\ (\Delta T)_m}$$

where: θ = drying time, hrs.

ρ_s = particle density, lb. / cu. ft.

D_p = $\sqrt{A_p/3.14}$ effective particle diameter, ft.

A_p = area of particle, sq. ft.

λ = latent heat of vaporization, B.T.U./lb.

W_o = initial particle moisture content, $\frac{\text{lb. } H_2O}{\text{lb. B. D. mat'l}}$

W_1 = final particle moisture content, $\frac{\text{lb. } H_2O}{\text{lb. B.D. mat'l}}$

k = thermal conductivity of air film, B.T.U./hr.-sq. ft.-^{o} F./ ft.

$(\Delta T)_m$ = average temperature difference between inlet and outlet conditions, ^{o} F.

Subscript = average condition of gas film

The length of the drying section is then given by the product of the gas velocity times the drying time per particle.

$$L = V\theta$$

where: L = length of drying section, ft.

V = velocity of gas stream, ft./hr.

This procedure applies only to the case where the wet feed is dispersed by the hot gas stream in a constant throughput drier.

The velocity of the conveying gases must be high enough to transport the largest particle to be dried. This velocity must then be in excess of the velocity required to just float the particle in the gas stream. Perry (6) advises that in actual practice the velocity for pneumatic conveying drying should be about 100 percent greater than the calculated free-

falling velocity for the largest particle. The velocity then in many instances would range as high as 200 feet per second. The quantities of air required are also dependent upon other factors such as the amount of moisture to be removed, the available temperature drop of the air, and the quantity of the solid material to be conveyed. Perry (6) also states that the ratio of solids to conveying gas normally ranges from 0.05 to 1.0 lb. solid per lb. of conveying gas.

The mass- and heat-transfer rates in a flash drier depend upon the gas rate, the gas stream temperature, the humidity of the gas stream, the particle size, and the particle shape. The particle size, shape, and density fixes the mass velocity required to float the particle in the rising gas stream. As the mass of a given size particle is decreased, a lower mass velocity is required to support it and the flow past the particle is decreased. The chief effect of the velocity past a particle was on the mass- and heat-transfer coefficients. The flow of gas at high velocities past the particles reduces the thickness of the vapor film surrounding the particle. A still further increase in the velocity past the particle further decreases the film thickness, and in turn, the resistance to diffusion.

Considerable work has been done to determine the effects of mass velocity past a stationary surface with respect to a given position of the surface in the flow path. None of these correlations apply to flash drying because the particles floating in the rising turbulent gas stream are continually moving

about. Therefore, the velocity past the surfaces of the same particle will not be equal and each will display different rates of mass and heat transfer.

High temperatures are essential to rapid rates of drying as well as to high termal efficiency because the greater the differential temperature between the gas stream and the surface of evaporation the greater is the driving force to cause heat transfer through the gas film. The temperature of the particle surface is taken as the wet-bulb temperature of the gas stream. The difference between the vapor pressure of the water vapor at the surface of the particle and the partial pressure of the water vapor in the main gas stream constitutes the driving force for the mass transfer of water vapor through the gas film. Any change in conditions that bring about an increased spread between the dew point and the wet-bulb temperature increases this driving force. One of the most common methods of spreading this range is to increase the dry-bulb temperature of the gas while holding the dew point constant; then as the dry-bulb temperature is increased the wet-bulb or saturation temperature will also be increased. Any change in conditions that bring about an increase in the temperature differential will also bring about a corresponding increase in the vapor pressure differential. Thus the rates of mass and heat transfer are dependent upon each other.

EXPERIMENTAL STUDIES

Theoretical Approach

In the past few years considerable work has been done with j transfer factors which were first developed by Colburn (1) for the gas film. Hougen and co-workers (3,7,8) report a similar correlation for mass and heat transfer through the gas film for data obtained on porous ceramic spheres, cylinders, and commercial tower packings such as raschig rings, partition rings, and Berl saddles. These data were obtained by measurement of the rate of evaporation of water into an air stream from wetted granules and packings during the constant-rate period (3,7,8). The investigations carried on by Gamson, Thodos, and Hougen, Taecker and Hougen, and Wilke and Hougen are analogous to the drying that takes place in a flash drier if three limiting assumptions are made. They are:

1. The dispersed particles in the conveying gas remain at all times in a constant relative position with respect to each other.

2. Equal dispersion of all particles in the gas stream and thorough mixing of all gases.

3. The heat available for evaporation of water comes only from the gas that flows past the particles.

These assumptions allowed the consideration of the flash drying process as if it took place in a fixed bed of extremely large void fraction, and as if the gas rate through the bed would be

the gas rate required to float the particle in the gas stream.

By consideration of the drying process in the above manner, it would seem possible that there should be a correlation between the mass- and heat-transfer factors for flash drying and those previously obtained for evaporation of water from the surface of particles in granular beds during the constant-rate period where the temperature of the evaporating surface is taken as equal to the wet-bulb temperature of the gas stream. Although the basic transfer coefficients, k_G and h_G, are more convenient to use for design purposes, they have the disadvantage of varying widely with changes in mass velocity, gas stream temperature, humidity, and particle size. Therefore, a correlation of j transfer factors as a function of the modified Reynolds number for flash drying processes would lend itself readily to the design of future driers.

Colburn (1) and Chilton and Colburn (2) introduced the j transfer factors to facilitate the plotting of experimental data for widely varying systems and conditions. These factors are combinations of the variables combined in dimensionless groups to define what is termed a mass-transfer factor, j_d, and a heat-transfer factor, j_h. They are as follows:

$$j_d = \left(\frac{k_G \; M_m \; P_f}{G}\right) \left(\frac{\mu}{\rho D_v}\right)_f^{2/3}$$

$$j_h = \left(\frac{h_G}{C_p \; G}\right) \left(\frac{C_p \mu}{k}\right)_f^{2/3}$$

where: k_G = mass-transfer coefficient of gas film, lb. moles/hr. -sq. ft.-atm.

M_m = mean molecular weight of gas stream

P_f = film pressure factor, atm.

G = mass velocity, lb. /hr.-sq. ft.

μ = viscosity of gas, lb./ft.-hr.

ρ = density of gas stream, lb./cu. ft.

D_v = average diffusivity coefficient of water vapor moistures, sq. ft./hr.

$(\mu / \rho D_v)$ = Schmidt number, dimensionless

h_G = heat-transfer coefficient of gas film, B.T.U./ hr.-sq. ft.- ^{o} F.

$(C_p \mu / k)$ = Prandtl number, dimensionless

Subscript $_f$ = properties at average condition of gas film

The evaluation of the transfer factors can be made from experimental data relating the temperatures, the dew point of the gas stream, or both, for all points in the drier. In this investigation both temperature and dew point data were obtained.

The amount of water evaporated per hour from the feed material can be calculated as the product of the specific heat, temperature drop, and mass velocity through the drier divided by the latent heat of vaporization of water at the wet-bulb temperature of the entering gas stream. The water evaporated per hour from the feed material, based on dew point data, can be taken as merely the change in moisture content of the gas stream in pounds of bone dry air per hour. The number of pounds of water vaporized per hour over the molecular weight of water and the surface area of the drier hold-up gives r_{Aa}, the molal rate of diffusion of water vapor through the gas film. Then through use of the equation:

$$r_{Aa} = k_G (\Delta P)_m$$

where: r_{Aa} = molal rate of diffusion, lb. moles/hr.-sq. ft.

$(\Delta P)_m$ = Log mean pressure difference of water vapor, atm.

the mass-transfer coefficient for the diffusion of water vapor through the controlling gas film can be calculated. The value of the log mean pressure difference was based on the water vapor pressures at the terminal conditions. The mass-transfer coefficient calculated would then represent the average condition of the run.

Evaluation of the heat-transfer coefficient of the gas film could be done in a similar manner. The heat required to vaporize the water and the cross-sectional area of transfer are known from the calculation of the mass-transfer coefficient. This leaves only the log mean temperature driving force to be determined from experimental data so that the equation:

$$Q = h_G A (\Delta T)_m$$

where: Q = heat to vaporize water, B.T.U./hr.

could be used to calculate the value of the heat-transfer coefficient. The values of k_G and h_G can then be used to evaluate the j factors for mass and heat transfer through the gas film for the average conditions of each run.

Equipment

Furnace. Hot gas for the drying system was supplied by the combustion of natural gas in a hot air furnace. Natural gas at a line pressure of 5 pounds per square inch and air at 100 pounds per square inch were passed through regulating valves and then to a mixing nozzle where they were mixed and

ignited. The products were then passed into a combustion tube where any unburned gas was mixed with air and ignited when it came in contact with the hot checkerwork. In order to insure maximum mixing of the partially burned gas and air, the combustion tube was constructed in the shape of a venturi, having a throat diameter of 7 inches increasing to 1 foot square over a distance of 4 feet. The furnace proper was 9 feet long, 4 feet wide, and 3 feet high. Both the furnace and the combustion tube were constructed of fire brick.

Since the combustion gases were only a small part of the gas required by the drier, air was drawn through the front of the furnace and allowed to mix with the combustion gas to obtain a mixture of the desired temperature. On the average the warming up period required 20 minutes before a constant inlet gas stream temperature was attained,

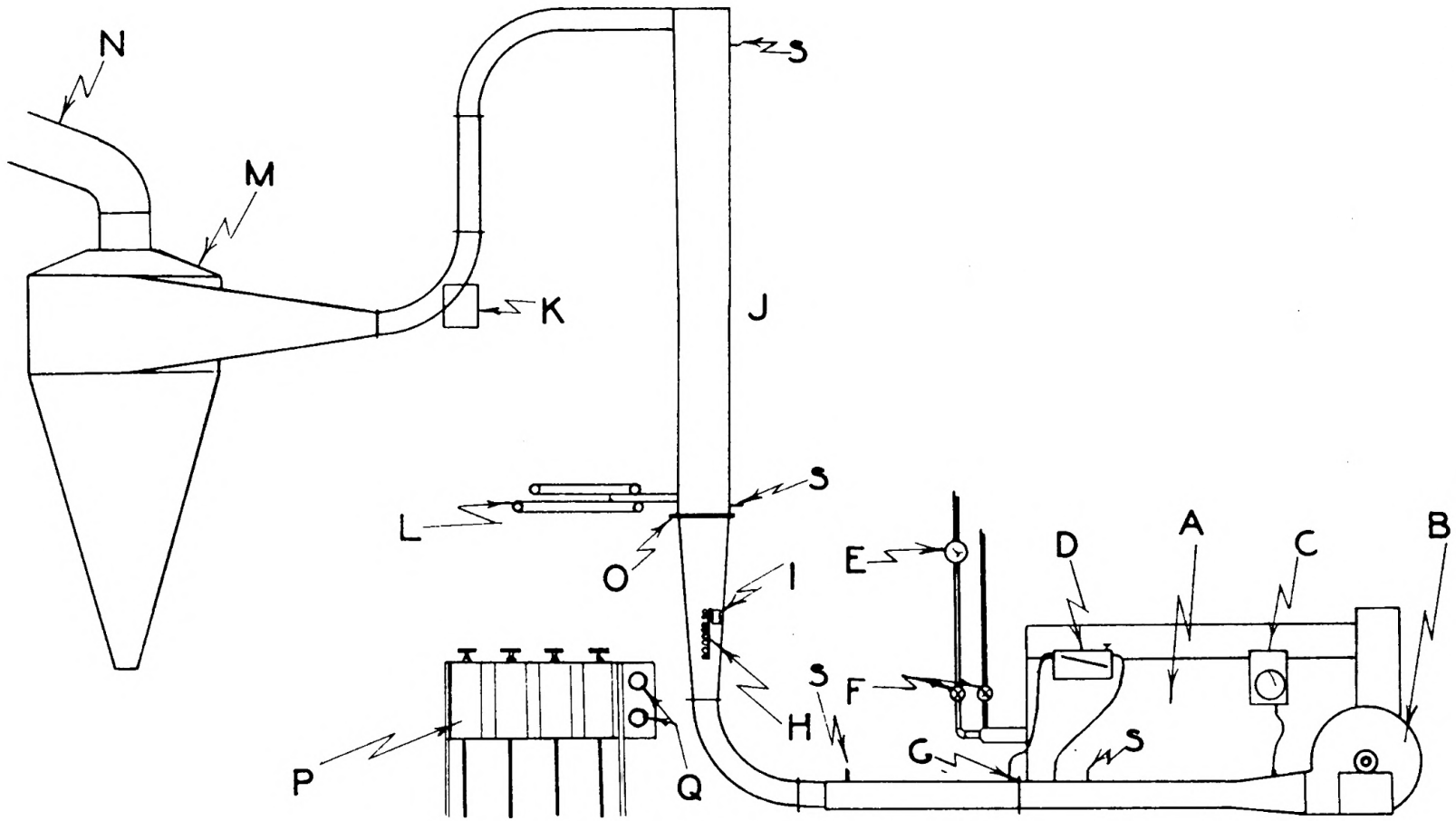
Blower. Circulation of the hot gas through the drier was accomplished through the use of a Clarage No. 15 exhauster fan, equipped with long shaving blades, powered by a 10 horse-power Allis-Charmers induction motor. The coupling between the motor and the fan consisted of four 1-inch V-belts over two 6-inch compound pulleys. The suction side of the fan was connected to the back of the furnace by a conical transition section. This section was constructed of 24-gauge sheet iron, 2 feet in length, with a furnace entrance 8 inches in diameter, and a blower suction end 15 inches in diameter. Means for the regulation of the mass flow-rate was provided by a slot on the intake side of the blower into

EXPLANATION OF PLATE I

Pneumatic Flash Drier

A. Hot air furnace
B. Clarage no. 15 exhauster fan
C. Wheelco temperature recorder
D. Inclined manometer
E. Gas pressure gauge
F. Gas and air control valves
G. Orifice plate
H. High velocity thermocouple tubes leaving drier
I. Jones plug box for thermocouple leads
J. Drying section
K. Hold-up trap
L. Injection feeder
M. Cyclone separator
N. Exhaust gas duct to flue
O. Support flange
P. High temperature dew point analyzing equipment
Q. Variable voltage transformers to supply analyzer heater potential
S. Static pressure taps

PLATE I



which orifice plates could be inserted. These orifice plates had openings of 3-7/8, 4-1/2, 5-1/8, 6, 7-7/8 and 10 inches in diameter. When no plate was used the opening was that of the intake of the fan. Discharge from the fan was into another transition section, 13 inches square at the blower and reduced to 8 inches in diameter over a length of 4 feet. Connection to the drier was made with 14 feet of 8-inch sheet metal duct and a 90° elbow which had an inside radius of 2-1/2 feet. This elbow was constructed of sheet metal and had a 7 inch square cross-section. Connections between the elbow and the duct and the elbow and the drying section were made with transition pieces, 10 inches long, going from 7 inches square to 8 inches in diameter.

Drying Section. A transition section, 67 inches long expanded from 8 to 16 inches in diameter, was used to connect the elbow to the lower end of the drying section by means of a flange. The drying section consisted of a vertically suspended metal duct, 6 feet in length and 16 inches in diameter. Above this was an 8-foot 4-inch transition piece which gradually increased the column diameter from 16 to 18 inches. The upper end of this transition piece was closed to form the top of the drier.

Rising through the center of the column was a bundle of three-eights inch O.D. copper tubes. These tubes had a twe-fold purpose. First, they acted as high velocity thermocouples and secondly, they provided a means for sampling the gas stream at various levels throughout the column.

The inlets to these tubes were so arranged that they facilitated the measurement of the temperatures and taking of samples of the gas stream at one foot intervals starting 12 inches from the top of the drier. There were 16 of these tubes which extended vertically downward from their inlet level into the lower transition section where they were brought out of the drier, 8 on each side. At a point 13 inches below the flanged joint was a diverting cone made of 14 mesh stainless steel screen. This cone prevented the feed material from falling lower in the system and tended to upset the fall of the particles that slid edgewise down the wall of the drier. By tipping the discs so that their broad side was exposed to the rising gas stream they floated upward again. The feed entered the drier at a point 5 inches above the support flange and the dry product was taken off the top in an 8 inch duct.

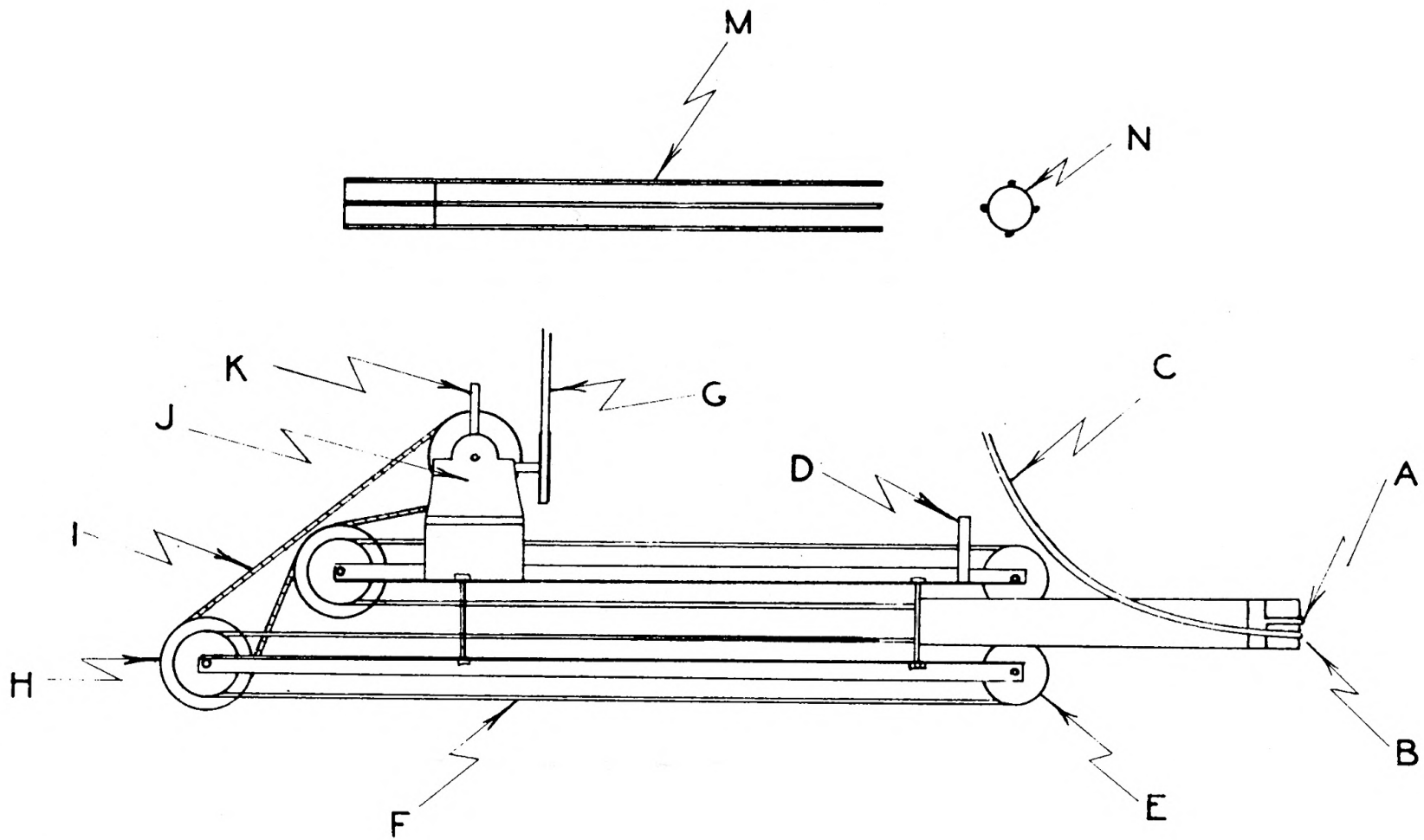
Feed. The feed material used in this investigation was unwaxed, number 9, milk bottle caps. These caps had a diameter of approximately 1-5/16 inches. The surface areas of the caps remained constant throughout the drying period. This constant surface area per particle was essential in order to properly evaluate the transfer rates. The moisture content of the wet feed was adjusted by the period of time the caps were soaked before placing in the humidor and the number of previous soaking and drying cycles. As the soaking period or number of wettings increased the caps swelled to about twice

EXPLANATION OF PLATE II

Injection Feed System

A. Drag claw
B. Air jet
C. Air line to jet
D. Front injector support
E. Two inch pulley
F. One-half inch V-belt
G. V-belt drive from Reeves variable speed drive
H. Chain drive sprocket
I. Chain drive
J. 50-1 gear reducer
K. Rear injector support
M. Feed cartridge
N. End view feed cartridge

PLATE II



their initial thickness with no appreciable change in diameter. This swelling gave an increased porous volume in which water was held.

Feeder. The feeder was of the continuous injection type designed to give a constant feed rate. Basically it consisted of four 1/2-inch V-belts, each stretched tightly over a pair of 2 inch pulleys. Each pair of belts and pulleys was mounted on a common shaft so that the inner edges of the belts were in parallel planes 1 inch apart. The 2 banks of belts were brought together so that the inner edges of the belts gripped the feed disc at intervals of 90 degrees around its circumference. In order to insure a constant forward movement of all belts, the two banks were provided with a chain drive from a 50 to 1 gear reducer mounted directly above the supporting frame. The power to drive the feeder was supplied from a Reeves variable speed drive. Feed discs were fed into the back end of the belt conveyer where they were gripped by the belts and then carried forward at a constant rate to be discharged into the injection tube. The injection tube had an inner diameter of 1-7/16 inches and was 1 foot long. At the discharge end of the tube was located a pair of drag claws, one on each side, and an air jet. The function of these was to break up the cap slug being discharged from the tube and allow them to enter the gas stream one at a time.

Since the belt speed for a feed rate of 120 pounds per hour was 6-1/2 feet per minute, the discs were inserted into

the feed end of the conveyer from feed cartridges, each containing approximately a 1 foot slug of discs. This cartridge was removed as soon as all the caps were gripped by the belts and then a new cartridge was inserted. The cartridges were merely a 4 pronged carrier built to hold a slug of discs approximately 12 inches long. The discs were held loosely by the prongs which were located at 90 degree positions around the circumference of the discs. When inserted into the feeder these prongs slipped between the belts allowing the belts to grip the caps.

Cyclone Separator. In order to separate the dried product from the hot gas stream a cyclone separator was included in the system. This separator was not designed for use with this drier but was utilized because it was available. This investigation was primarily interested in the drying section alone and not in the method of separating the product. This separator in overall length was 11-1/2 feet; the conical section was 9 feet in length, and 6 feet in diameter at the top reducing to 7 inches at the bottom. Above the conical section was located a cylindrical section 2-1/2 feet high. Out of the top of the separator an 18-inch 30-foot long duct carried the exhaust gases to the chimney where they were released to the atmosphere. The separator was connected to the drier with 5 feet of 8-inch diameter duct and 2 elbows similar to the one at the base of the drying section.

Hold-Up Trap. Knowledge of the hold-up in the drier for any given set of operating conditions was essential; therefore, a hold-up trap was included at the lower elbow just before the hot conveying gas entered the separator. The device consisted of a trap-door in the outer radius of the elbow, provided with a system by which the trap could be tripped inward so as to divert the gas stream from the separator into a large wire basket when the feeding was stopped. Any material remaining in the drier at this time was retained in the basket. This wire basket was made of 16 mesh galvanized screen wire, 7 inches wide, 14 inches deep, and 18 inches high.

Instrumentation

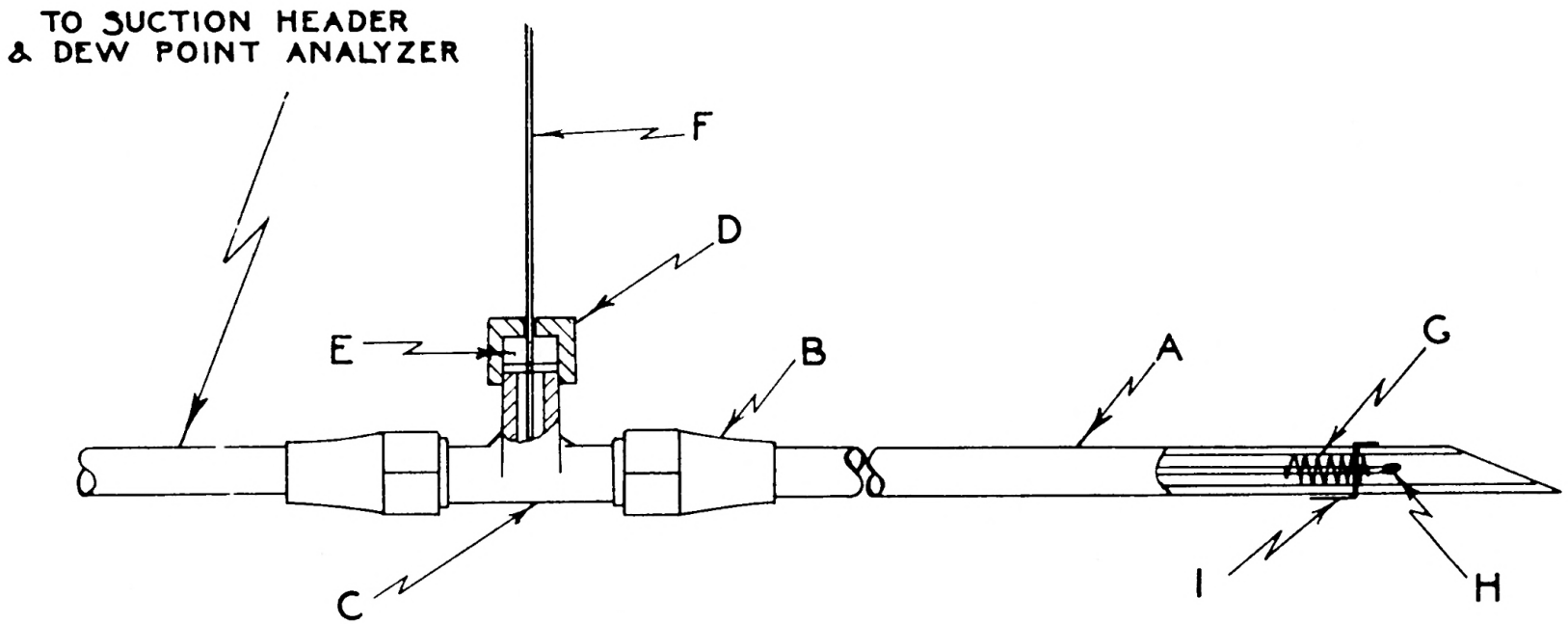
Instrumentation and Control. The inlet gas stream temperature was manually controlled by means of the gas and air regulating valves with the help of a Wheelco temperature recorder which had its sensitive element located in the duct, one foot from the blower. A 7 inch sharp edge orifice was located at a distance of 8-1/2 feet from the blower with its radius taps located 8 and 4 inches up and down stream respectively from the orifice. These taps were connected to a Merriam inclined manometer with heavy walled rubber tubing. Also located in the system were four static pressure taps each connected to a U-tube manometer which had one leg open to the atmosphere. Two of these taps were located on the drier. One tap was opposite the feed port, 4 inches above the support flange, and the other was 1 foot from the top of the drier.

EXPLANATION OF PLATE III

High Velocity Thermocouple Detail

A. Three-eights inch O.D. copper tube
B. Standard 3/8 inch brass union nut
C. Standard brass T, 3/8 inch tube to 1/4 inch pipe to 3/8 inch tube
D. Standard 1/4 inch pipe cap
E. Furnace cement plug
F. Double strand fiberglas covered iron constantan thermocouple wire
G. Spiral support and aligning spring
H. Welded thermocouple junction
I. Support pin for spring

PLATE III



The other 2 static pressure taps were located on the connecting duct between the blower and the drier, one 20 inches up stream and the other 5 feet 4-inches down stream from the orifice. All manometers were filled with distilled water.

High Velocity Thermocouples. As mentioned before, the temperatures of the gas stream at various levels in the drying column were measured through the use of high velocity thermocouples. The couples for the most part were constructed of standard fittings. The inlet ends of the tubes were beveled so that when the tubes were mounted in a vertical position in the center of the drying section there would be no shelf or obstruction formed onto which discs slipping or falling down the tube bundle could become lodged.

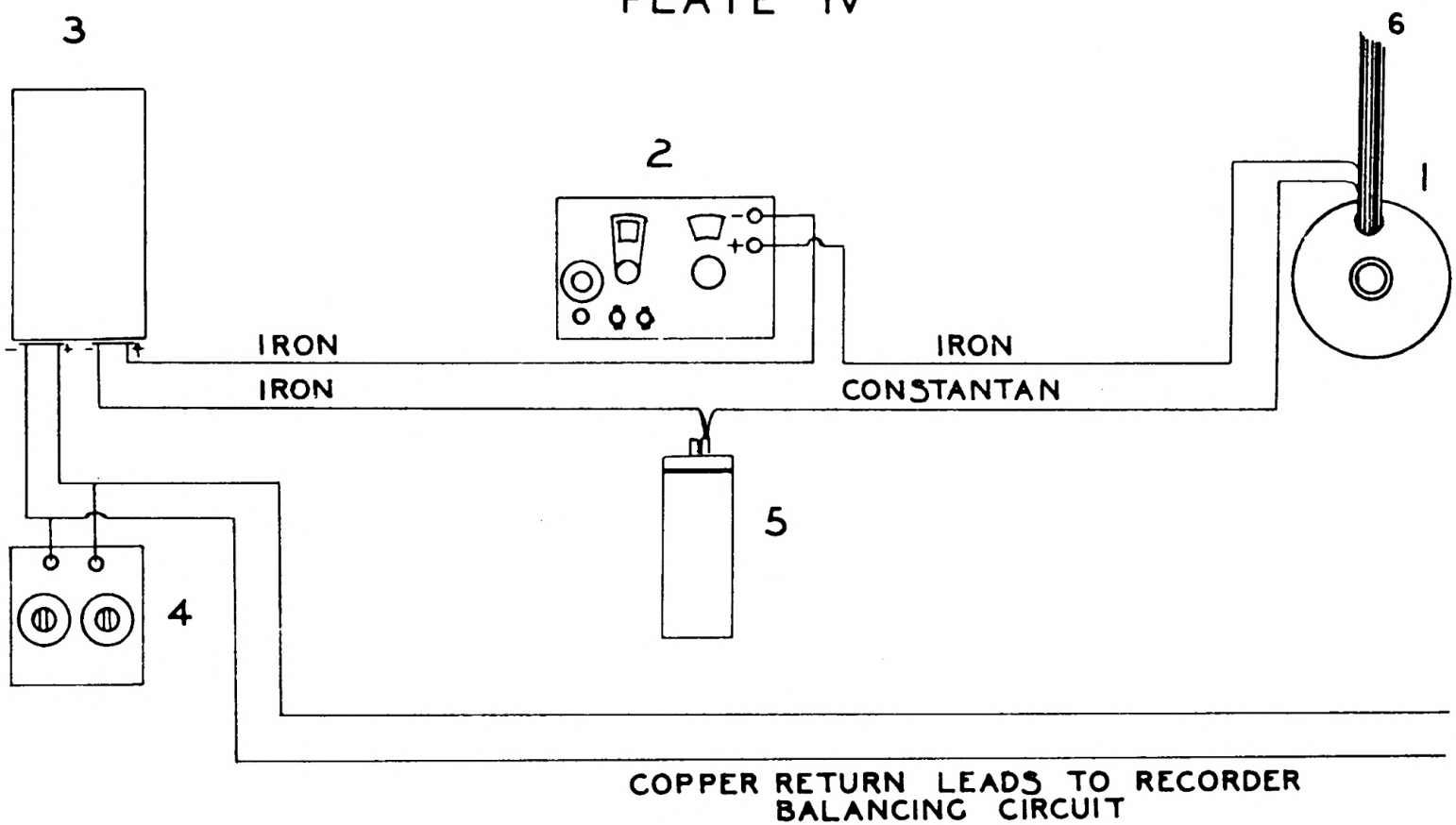
The length of the tubes, of course, depended upon the positions of their inlets in the column. Couples located high in the column had longer suction tubes than those located at lower positions. Since the length of these tubes was so short, varying pressure drops were considered negligible and the flow of gas through each was substantially equal. Any slight variation in flow between the tubes, however, had no effect on the operation of the couple because the primary objective was to get a high velocity past the junction for a minimum of gas flow. There were two decided advantages in using high velocity thermocouples to measure these temperatures. First, the high velocity of the gas past the couple junction decreased the gas film thickness at the junction, thus increasing its sensitivity

EXPLANATION OF PLATE IV

Recorder Range Spreading Circuit

1. Selector switch in recording potentiometer
2. Potentiometer to supply bucking potential
3. General Electric Self Balancing Potentiometer
4. Decade resistance box
5. Thermos cold junction

PLATE IV


3
2
6
1
IRON
IRON
IRON
CONSTANTAN
5
4
COPPER RETURN LEADS TO RECORDER
BALANCING CIRCUIT

to changes in temperature; secondly, passage of the hot gas back along the couple leads reduced to a minimum the heat loss from the junction by conduction along the leads. The thermocouple leads were brought out of the suction tube just after the tubes emerged from the lower part of the drying section, Plate III. Then for convenience in erection, the leads went to multi-terminal Jones type plugs located on each side of the drier.

Drying Section Temperature Measurement. The temperatures measured by the high velocity couples were recorded by a Brown, 16-point, Electronic recording potentiometer. Initially this instrument was constructed to measure temperatures in degrees Fahrenheit from 0-800 degrees for iron-constantan thermocouples. To better suit the needs of this investigation the instrument was converted to read 0-100 millivolts using an external cold junction of 32° F. By doing this and using the circuit shown in Plate IV, it was possible to expand any portion of the 100 millivolt range of the instrument to a full-scale reading.

This feature greatly increased the accuracy of the temperature measurement over the narrow range of temperature drops throughout the column. The lower terminal of the range was determined by the bucking potential applied on the circuit. This potential was then indicated as zero on the recorder scale. From the potentiometer that supplied the bucking potential the difference in potential between the thermocouple and the bucking potential was fed to a General Electric self-balancing potentiometer. Since the hot junction was located in the

column and the reference junction was in the ice bath, all connections except the one from the selector switch to the cold junction were made with iron wire.

The General Electric self-balancing potentiometer was designed to measure potential differences in the microvolt range. This instrument was very useful as a device for use in conjunction with a recording potentiometer of fixed range, because the self-balancing potentiometer was useful as both a direct current voltage amplifier or voltage divider; therefore, any desired recording potentiometer range was possible.

Since the output of the General Electric potentiometer was in microamperes instead of microvolts, as required for operation of the Brown multi-point recorder, and to facilitate the spreading of the range, a variable resistance or decade box was inserted across the output terminals of the General Electric instrument.

Ohm's law states that $E = IR$, therefore, for any given current, I, an increase in resistance, increased the resulting output voltage, E. Likewise, a decrease in the shunt resistance resulted in a decreased output voltage for the same applied current. Then to spread the scale reading for a given potential difference between the couple junction and the applied bucking potential merely required the proper adjustment of the shunt resistance.

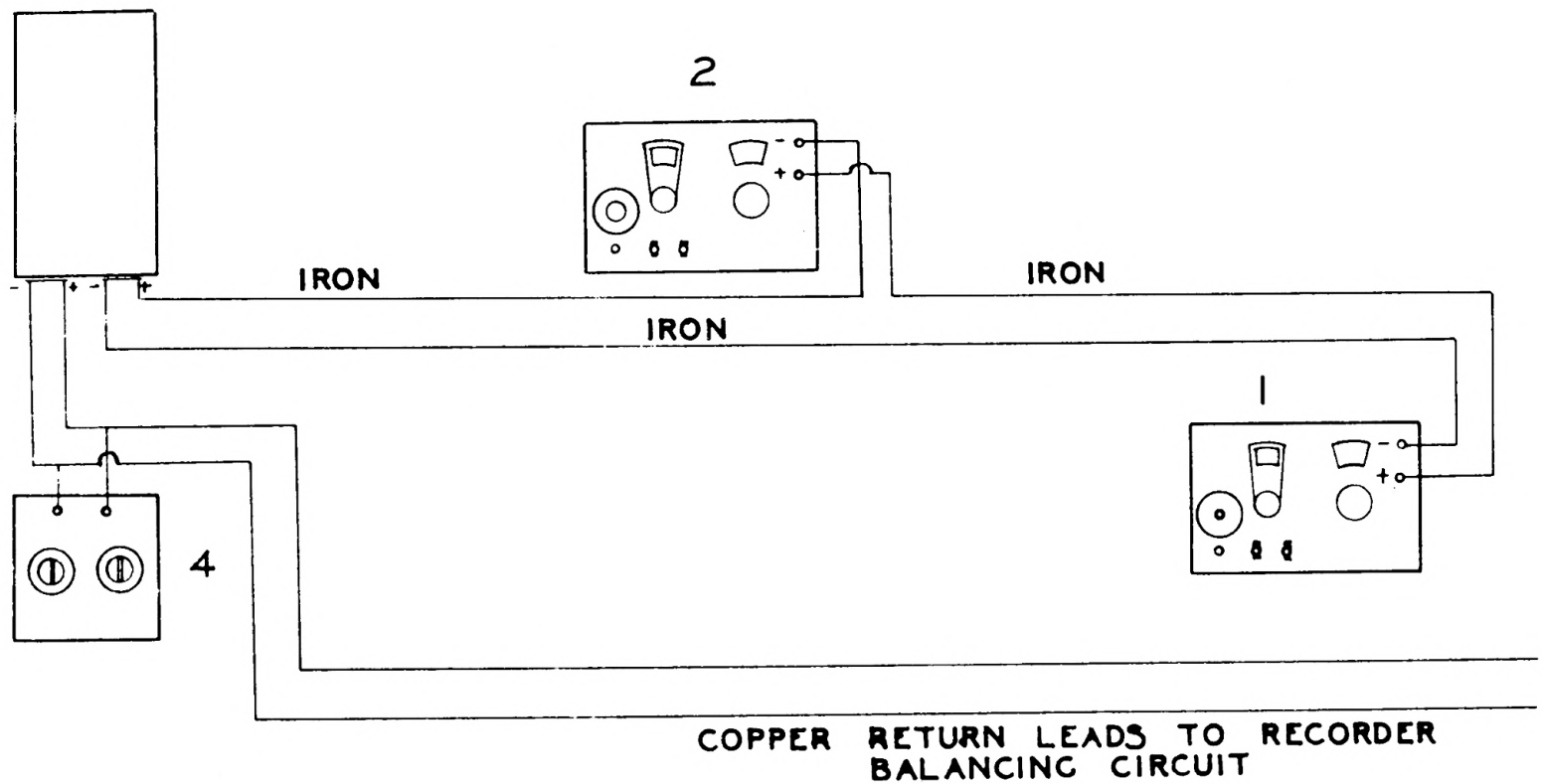
Calibration of any predetermined range was carried out by inserting a potentiometer in the circuit as shown in Plate V for the thermocouples. For example, it was desired to measure

EXPLANATION OF PLATE V

Recorder Calibration Circuit

1. Calibration potentiometer
2. Potentiometer to supply bucking potential
3. General Electric self-balancing potentiometer
4. Shunt resistance decade box

PLATE V


3
2
IRON
IRON
IRON
1
4
COPPER RETURN LEADS TO RECORDER
BALANCING CIRCUIT

temperatures corresponding to potentials falling in the range of 5 to 13 millivolts. Since the lower limit was 5 millivolts, this potential was applied with a portable potentiometer as the bucking potential. The maximum potential was 13 millivolts and was applied to the circuit by the potentiometer temporarily substituted for the thermocouples. Adjustment of the shunt resistance then allowed the 8 millivolt difference to be spread over any desired portion of the scale length. Intermittent points in the selected range were recorded by merely transgressing the range at equal intervals from 5 millivolts to 13 millivolts with the substituted potentiometer. The lower 5 millivolt reading was the zero scale reading and the chosen intervals were then distributed equally between zero and the set maximum scale reading.

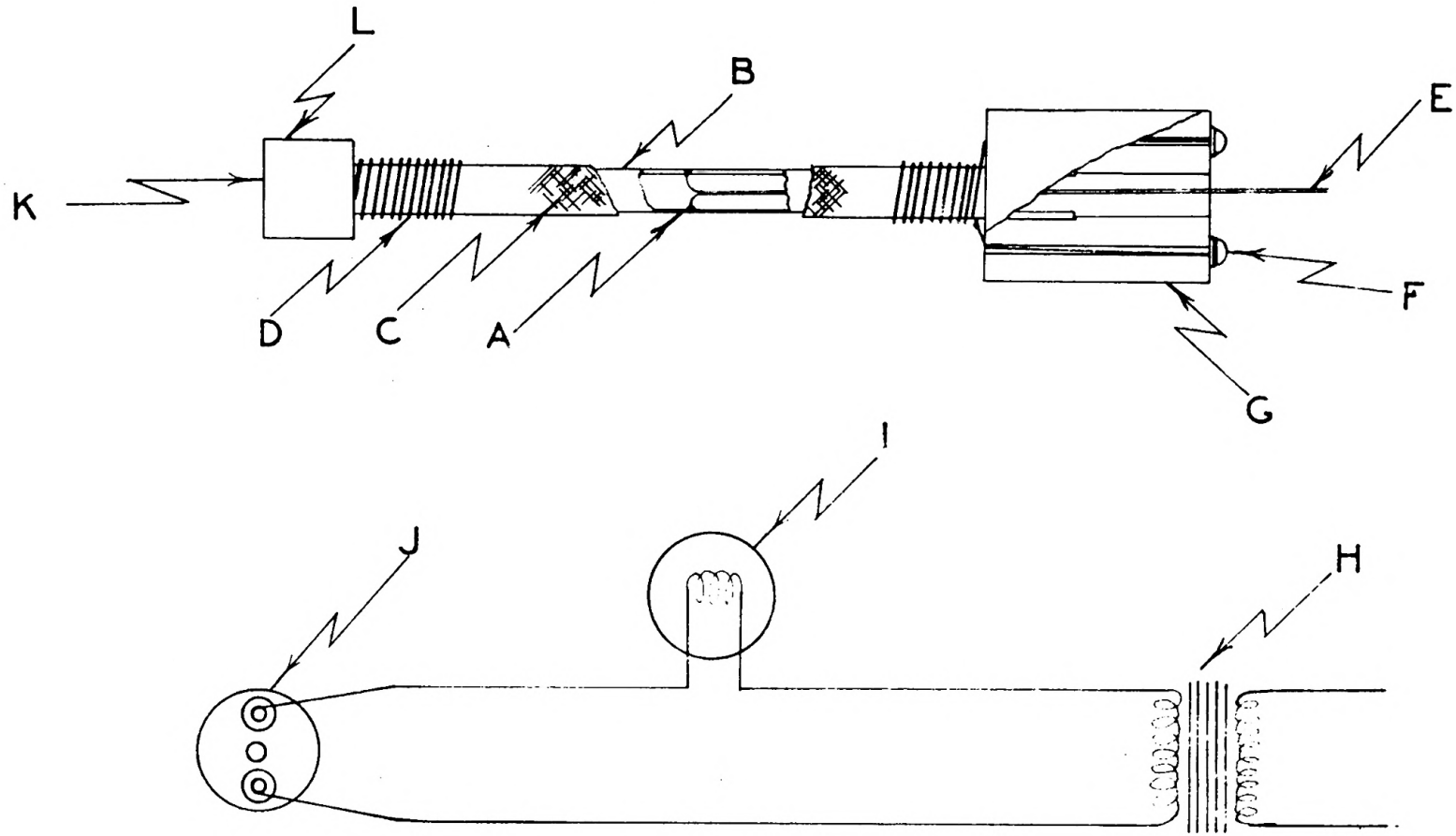
High Temperature Dew Point Equipment. The gases drawn through the high velocity thermocouple tubes, after leaving the column, passed either directly to a low vacuum suction header or to the intake ports of the high temperature dew point analyzers. Unless the gas samples were being taken the gas passed up through a ball check valve and then to the low vacuum suction header. This low vacuum header had the sole purpose of supplying flow through the high velocity tubes when no samples were being taken. The ball check valve was inserted in the line to prevent reversal of flow in the system when the sample was drawn into the dew point analyzing equipment.

EXPLANATION OF PLATE VI

Analyzer Unit Detail

A. Soldered thermocouple junction
B. Varnished brass tube
C. "Fiberglas" sleeve
D. Silver wire winding
E. Double strand fiberglas covered iron constantan thermocouple wire
F. Heater terminals
G. Plastic head
H. 25 volt output transformer
I. 25 volt 30 watt ballast lamp
J. End view analyzer head
K. Analyzer end cap
L. Phenol formaldehyde varnish coating

PLATE VI



The heart of the high temperature dew point analyzer was patterned after the Foxboro Dewcel. The analyzer consisted of a phenol formaldehyde varnished metal tube or socket for the temperature measuring element, a wick of woven "Fiberglas" sleeving around the tube, and 2 parallel silver wires wound around the wicking on approximately 1/32 inch centers. Anchorage for the windings was provided by plastic end caps. The wicking was impregnated with a 2 percent solution of lithium chloride, a hydroscopic salt. In operation the wires are connected in series to a 25-volt A-C power source with a 25-volt 30-watt standard light bulb which served as a ballast lamp. Imbedded in the varnished tube was an iron-constantan thermocouple that acted as the temperature sensitive element.

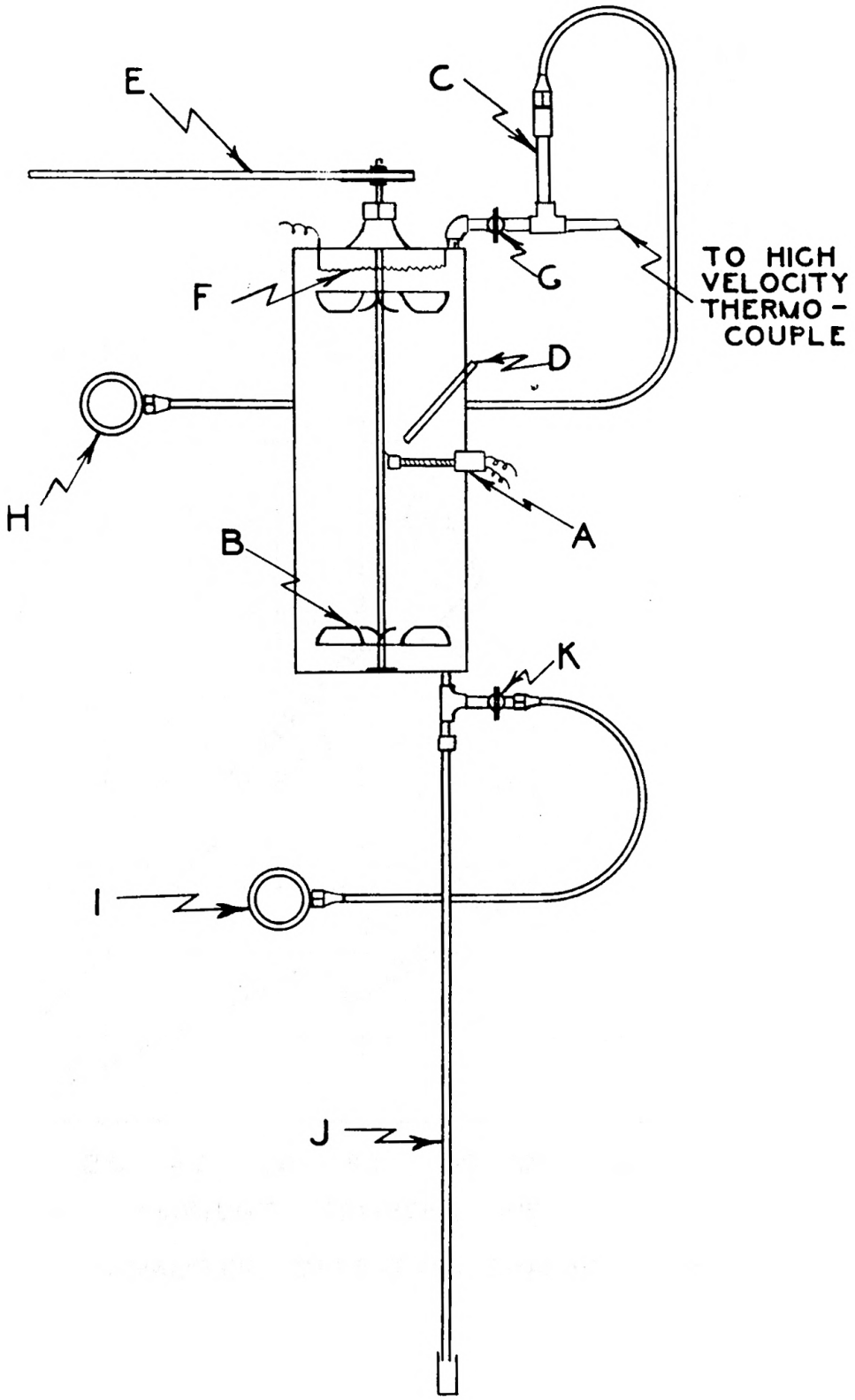
Moisture determination by the analyzer was based on the fact that for every water vapor pressure in contact with a saturated salt solution there is an equilibrium temperature at which the solution neither absorbs or gives up moisture to the surrounding atmosphere, Below this equilibrium temperature, the salt solution absorbs moisture. Above the equilibrium temperature the salt solution dried out until only salt crystals were left. The two electrodes, silver wires in contact with the wicking containing the salt solution, passed current only if the solution was at or below the equilibrium temperature when the salt was moist. The passage of electrical energy through the thin film of salt solution caused it to heat and quickly raised the temperature to the equilibrium point. Thus in effect the salt acted as an automatic self-regulator

EXPLANATION OF PLATE VII

High Temperature Dew Point Analysis Equipment

A. Analyzer unit
B. Circulation fan
C. Ball check valve
D. Thermometer well
E. V-belt drive for fan
F. Heating coil
G. Inlet valve
H. Low vacuum suction header
I. High vacuum suction header
J. Mercury manometer
K. Exhaust valve

PLATE VII



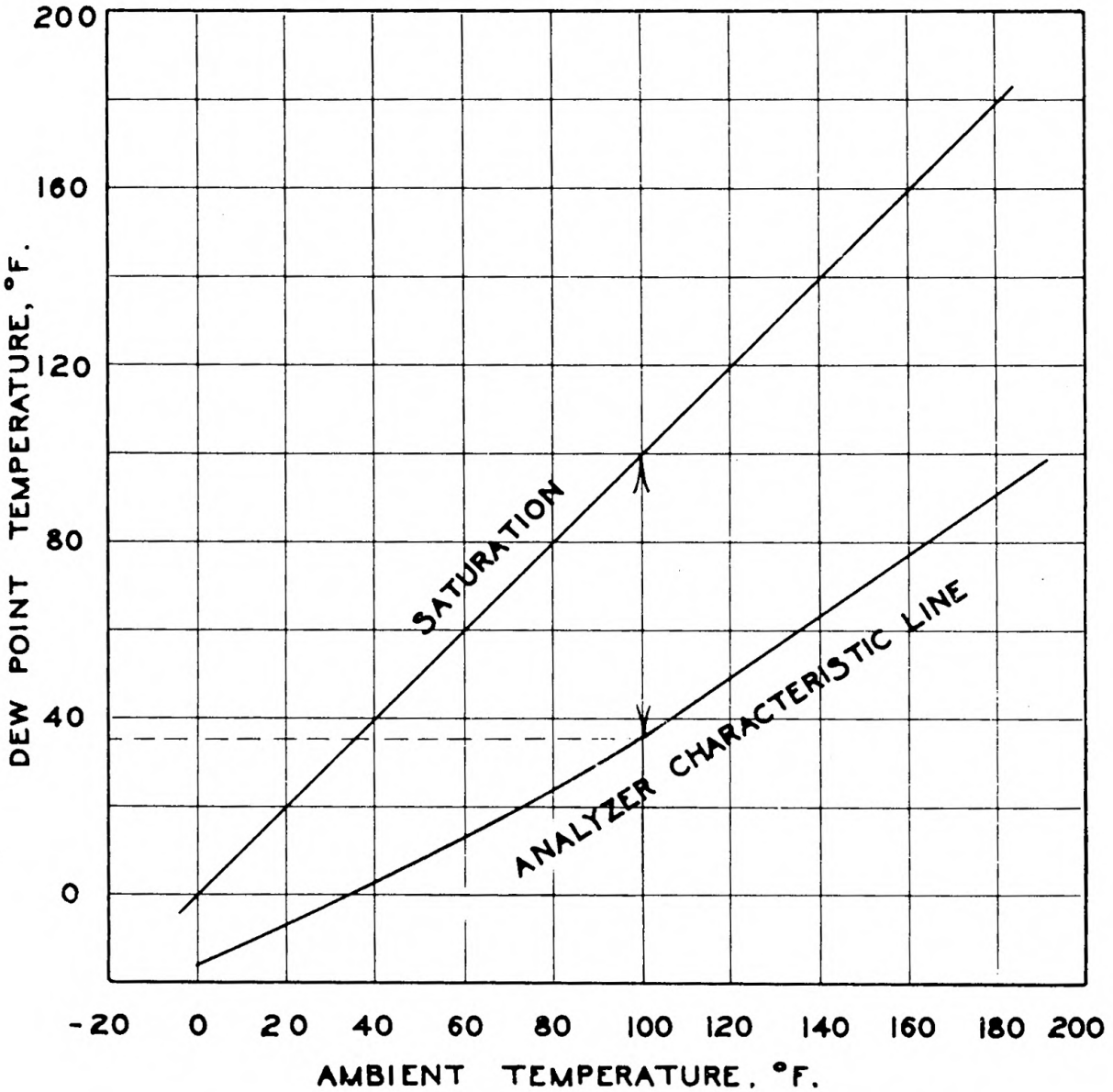


FIG. 1. ANALYZER OPERATING CHARACTERISTICS

to vary the flow of electrical energy in the heating system as required to maintain the salt at the equilibrium temperature. The equilibrium temperature, therefore, was a function of the absolute humidity of the ambient atmosphere.

The operating range of the analyzer is shown in Fig. 1. in terms of the dew point temperature and the ambient temperature. For example, at an ambient temperature of 100° F. dew point temperatures from 100° F. (saturation) down to 35° F. could be measured. If it was desired to measure dew points above 100° F., a higher ambient temperature was necessary. Therefore, the ambient temperature must at all times have been above the dew point temperature of the surrounding atmosphere or gas stream.

The high temperature dew point analyzing equipment were devices constructed for the sole purpose of maintaining the ambient temperature of the gas stream or sample within the operational range of the analyzer units. The high temperature dew point equipment consisted of a 5-gallon cylindrical, heavy walled container, provided with a fan for circulation, heater to maintain temperature, thermometer wall, analyzer unit, and intake and exhaust valves. Attached to the bottom of the container along with the exhaust suction control valve was a 30-inch mercury manometer. From the exhaust control valve a line lead to the high vacuum header. This high vacuum was necessary to sweep the unit free of previous samples and exhaust as much of the gas present as possible.

Materials and Experimental Methods

Preparation of the Feed. The unwaxed, number 9 (48 millimeter diameter) bottle caps or discs were soaked in water and then loaded into the feed cartridges 18 to 24 hours prior to each run. After being loaded the cartridges were placed in a large humidor where the entire mass of feed was allowed to come to an equilibrium moisture condition. On passage through the drier, especially during the low temperature runs, some of the caps were badly mutilated. These damaged caps were removed by hand sorting and replaced with new ones before the feed cartridges were reloaded. A few minutes prior to the run the humidor and its contents were weighed and placed at a convenient position with respect to the feeder.

Feeding. Approximately one minute before the run started the lid was removed from the humidor and the first caps were placed on the conveying belts of the feeder. To start the feeding, the power to the variable speed drive was turned on from a switch located close to the feeder. From the time the feeding was started until the completion of the run, the conveying belts were kept full of caps by merely inserting and then removing the empty cartridges when the caps were pulled from them. This required from 5 to 7 cartridges per minute, depending upon the length of the cap slug contained in each. The belt speed on the conveying system of the feeder was kept constant throughout the entire series of runs. After completion of each run the humidor, remaining

feed, and empty cartridges were again weighed to determine the actual feed rate.

Preparation Prior to Run. From one to two hours before each run was made the multi-point recorder was calibrated for the temperature range to be covered. This calibration was carried out as described in the previous section on equipment. After the calibration points were applied to the strip chart of the multi-point recorder the circuit was restored to that shown in Plate IV and the cold junction "Thermos" bottle filled with crushed ice. The recorder was now ready to accurately record the temperatures throughout the drier prior to and during the run.

Next the power to the dew point analyzers and heaters of the gas sample cans were turned on so as to aid in the removal of any remaining moisture from previous determinations. Fresh room air of known dew point was allowed to flow through the units by opening the inlet and then the exhaust cocks to the high vacuum header. This flushing was carried on for approximately 30 minutes. At the end of this flushing period the inlet cock was closed and as much as possible of the remaining air was evacuated from the gas sample cans. A high vacuum was maintained in the cans until just before the samples were drawn. At this time the heater potential on the cans was increased so as to maintain the analyzer temperature within the proper range for analysis.

The low vacuum used to keep a continuous flow through

the high velocity thermocouple tubes was now started and manually controlled to 4 inches of mercury vacuum. Next the multi-point recorder was started and the hot air furnace fired. Air to the jet on the feeder was also turned on and all points of the experimental apparatus at which data were to be taken were checked periodically during the warm-up period. The temperature of the hot gas stream entering the drier was controlled by the proper adjustment of the gas and air control valves so as to get the desired reading on the Wheelco temperature recorder. Once the desired entrance temperature was attained the system required about 20 minutes to reach equilibrium. As a rule no further adjustments were required to maintain the constant inlet gas temperature.

Sampling. Samples of the wet feed and product were collected every 2 minutes throughout the run. This was done by simultaneously removing 5 caps each from the feed and the product and placing them into the appropriate tightly covered sample weighing cans. The moisture contents were then determined by weighing the samples prior to and after drying. The samples were dried in a vacuum oven for a period of 20 hours at 100° C. and 25 inches of mercury vacuum. The feed sample was removed from one of the cartridges just before inserting it into the feeder and the product sample was taken from the separator. Except for the time when the addidional caps were placed in the sample cans these cans were kept tightly closed.

The "hold-up" in the column was determined by diverting the gas stream into the hold-up trap at the instant the feed was stopped. Flow into the trap was kept up until all of the caps had been blown from the column and collected in the basket. The hold-up was recorded in number of caps instead of weight because of the varying particle dryness after being subjected to the hot blast of air through the trap for varying lengths of time. In order to correct for the number of caps in the connecting duct between the trap and the drier, equal particle distribution was assumed for the drier and connecting duct. From the number of caps per unit volume the number of caps present in the connecting duct were calculated. Subtracting the number of caps in the connecting duct from the total collected gave the "hold-up."

Samples of the gas stream at eight different levels in the drier were taken from 7 to 10 minutes after the feeding was started. At this time during the run both thermal and mass equilibrium were attained. Just before drawing the gas samples the pressures in the evacuated gas sample cans was noted. One at a time the gas samples were drawn into the evacuated chambers by first closing the exhaust valve and then opening the inlet valve to allow the sample to be drawn in from the suction line. When the pressure in the chamber was equal to the pressure in the sample suction line the inlet valve was closed. This procedure was repeated until all of the samples were taken, the entire operation requiring around 2 minutes.

Analysis of the gas samples taken from the column was made one at a time. The first step in each analysis was to start the circulation fan so as to thoroughly mix the drawn sample with the air remaining in the partially evacuated vessel at the time the sample was drawn. Also it was necessary to provide circulation across the dew point analyzer unit.

Next the temperature of the dew point analyzer, which was a function of the dew point of the mixture, was taken at two minute intervals until 4 constant thermocouple potential readings were obtained on a potentiometer. Knowledge of the proper analysis temperature within the operational range of the dew point analyzer for the dew point of the gas sample came largely through experience gained in previous runs and in construction and testing of the analyzer units. As shown in Fig. 1, for an analysis temperature of 120° F., dew point temperatures from 48° F. to 120° F. could have been measured. With a range of almost 70° F. in dew point temperatures for a given analysis temperature, the most important requirement was to keep the temperatures in the gas sample cans at 10° F. to 15° F. above the maximum expected dew point.

Since it was impossible to thoroughly evacuate the chamber of the analyzer, the air remaining in the partially evacuated chamber was corrected for in the analysis calculations. When a chamber was partially evacuated to a known

vacuum the volume of the remaining gas referred to atmospheric pressure was equal to the total volume of the chamber times one minus the vacuum drawn on the chamber in atmospheres of pressure. Therefore, the volume of the gas sample in the chamber, if the final chamber pressure was one atmosphere pressure, was equal to the total chamber volume minus the volume of the unevacuated air. With the dew point temperatures of the initial air before evacuation of the cans and of the total mixture known, it was possible to calculate the dew point of the gas sample with the use of the dew point analyzer calibration curve. This dew point was the dew point of the gas stream in the drier at the level where the sample was taken. The same procedure was used on the remaining seven samples.

Recording of Data. The temperature at each point throughout the drying section was recorded on the strip chart of the multi-point recorder at approximately 20 second intervals prior to and during the run. Thirteen of the 16 temperature measurements made by the recorder were of the gas stream at various levels in the drier. The remaining 3 points were used to record the temperature of the drier wall at 3 selected positions. Once every 2 minutes during each run the following data were taken: pressure differential across the orifice, inlet gas stream temperature, the static pressures, natural gas and air pressures, and the weight of the product collected. The gas meter readings and the wet- and dry-bulb temperatures were recorded at the beginning and end of each run. The length

of each run and the weight of the material fed to the drier were recorded at the end of the run. Necessary data on the dew point determinations were taken during and after the run. The data were as follows: gas sample can temperatures, gas sample can pressures before and after drawing samples, and the dew point analyzer temperature in millivolts. The dew point analyzer thermocouple readings were made with a portable potentiometer with manual reference junction compensation.

INTERPRETATION OF DATA

General Discussion

Essentially there were four independent variables in each run: They were: gas rate, feed rate, inlet gas stream temperature, and the moisture content of the feed. Variables such as the product moisture content, gas stream temperature at the top of the drier, and hold-up in the drier were functions of the independent variables. Through the use of the j factors for mass and heat transfer, it was intended to relate the variables in such a manner as to facilitate the use of the correlation not only to the particular drier on which the data were obtained, but to aid in the design of other flash driers.

In the initial plan for the project it was desired to investigate the variables over wide ranges of inlet gas stream temperatures, mass velocities through the drier, and feed rates. During the first trial runs it was found, however,

that below a mass velocity of 2700 pounds per hour per square foot the dry product would not rise through the drying section. The feed settled in the drier until a layer or mat was built up on the diverting screen just below the feed port. This mat continued to build up until it almost completely blocked the flow of gas. At this time the entire mass would rise in the drier until the mat broke and the caps fell again to the bottom. This cyclic action continued until the drier was stopped, or if the temperature of the inlet gases to the column was high, the caps were ignited and burned as they became dry. This difficulty was expected as the experimentally determined support mass velocity for a single cap with its broad side exposed to the rising gas stream was 2850 pounds per hour per square foot. However, when the cap was turned edgewise in the stream it fell until something caused it to tip such that its broad surface was exposed again to pressure created by the motion of the hot gases.

The maximum temperature range for the experimental equipment was between room temperature and 500° F. Above an inlet temperature of 500° F., the critical mass velocity of 2700 pounds per hour per square foot was again reached. Essentially, the blower was designed to move a constant volume of air, therefore, as the temperature of the inlet gas increased the density per unit volume of entering gas was decreased and a lower mass velocity was obtained. If no restrictions were placed on the volume intake of the blower, the mass velocity in the drier was directly related to the gas stream tempera-

ture. Above 500° F. the blower was no longer capable of delivering the required gas mass velocity.

Data for runs 1 to 4 and 8 to 10 were not used in the correlation. Runs 1 to 4 had interruptions in the feeding due to clogging of the injection tube. This difficulty was overcome by rechecking the alignment of the caps in the feed cartridges and making any necessary corrections before placing them in the humidor. The data from runs 8 to 10 were discarded because of inconsistencies arising from measurement of the pressure drop through the orifice. Since the mass velocity was one of the primary independent variables, any error in its evaluation rendered the remaining data useless.

With the sensible heat necessary to carry on the evaporation of moisture from the particle available only from the gas which flowed past the particle, determination of the particle velocity through the drier was necessary. This was done with knowledge of the drier hold-up by calculating the number of particles contained in each foot high segment of the drying column. Then by division of the total number of particles per hour by the number of particles in each foot segment of the drying column, the velocity of particles in feet per hour was obtained. Subtracting this particle velocity from the gas velocity in feet per hour gave the effective velocity of the gas past the particle. The mass velocity past the particle based on the effective velocity is termed the modified or effective mass velocity of the gas stream. In order to better show the relationship between the actual and the modified mass velocity through the column, the corre-

lation was made using both velocities.

Correlation

The correlation of the variables investigated in this project are applicable only to the drying that takes place in the vertical drying section of the flash drier, since it was the principal piece of equipment under study.

Knowledge of the total particle surface area in a drier at any given time was essential before any further estimation of the mass and heat transfer rates can be made. In a flash drier this total surface area is dependent upon variables such as feed rate, particle shape, initial and final particle densities, and the mass velocity of the gas stream at the top of the drier. The mass velocity of the gas at the top of the drier controls the final density and in turn the final moisture content of the particle, assuming that no further drying takes place after the particle leaves the drying section. For any given particle size and essentially the same feed rate, the hold-up in the drying section was directly related to the outlet mass velocity of the drying gas stream. This relationship for the experimental data is shown in Fig. 2 where the hold-up in pounds of product per cubic foot of tower volume is plotted versus the outlet mass velocity of the gas stream. An increase in the mass velocity of the gas stream for any given feed rate resulted in a decrease in the hold-up. For the experimental data where the feed rate was approximately 124 lb. per hr. the hold-up for

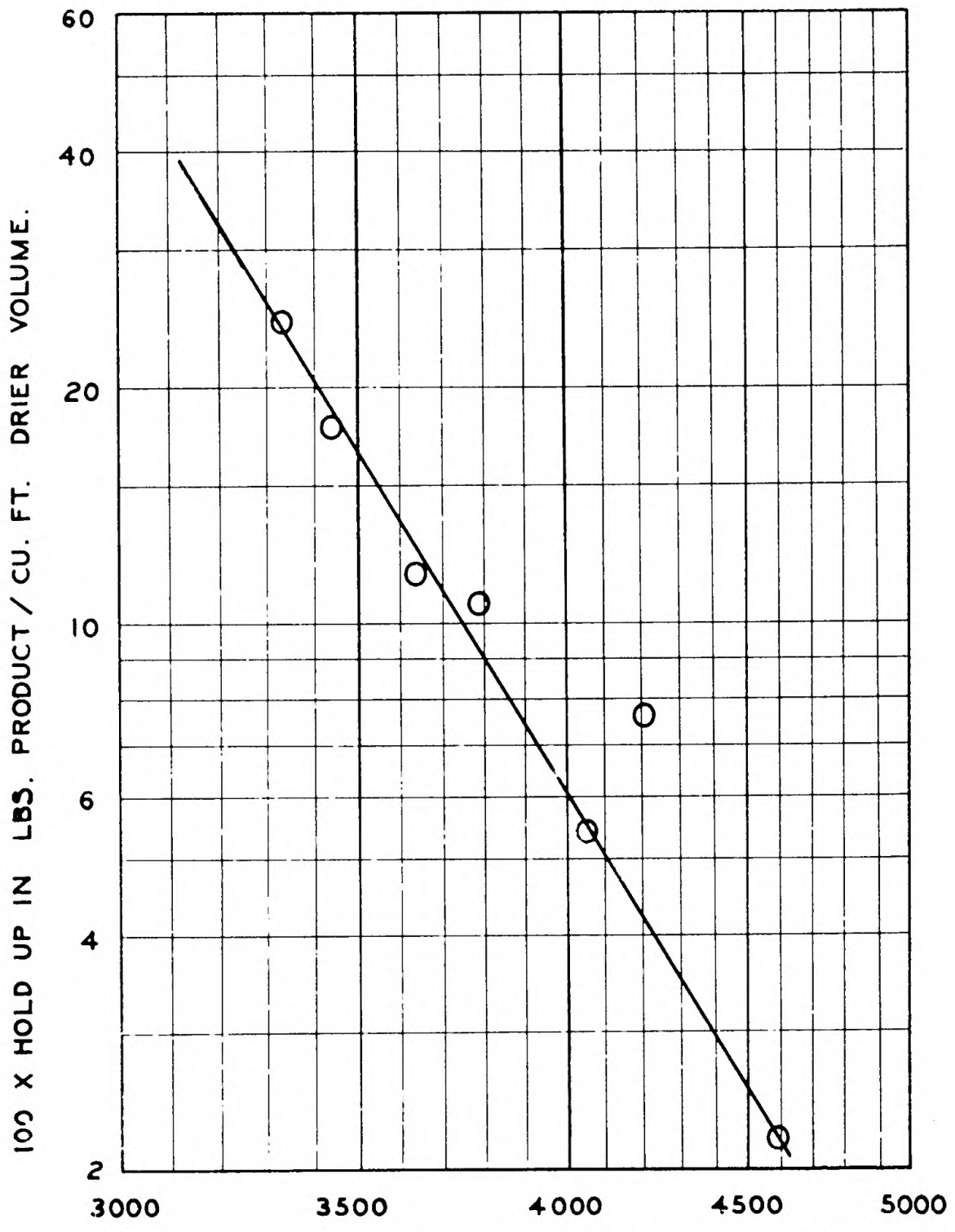


FIG. 2. DRIER HOLD UP VS OUTLET MASS VELOCITY

any given mass velocity is expressed as follows:

$$H = \frac{2.675 \times 10^{23}}{6.838^{G_2}}$$

where: H = hold-up, lb. product /cu. ft. drier volume

G_2 = outlet mass velocity, lb. /hr.-sq. ft.

With such a relationship existing between the hold-up in terms of pounds of dry product and the outlet mass velocity it was possible to relate the hold-up in terms of variables. Assuming that there was no change in the particle volume during the drying process, the ratio between the outlet and inlet density of the particle was expressed in terms of the ratio of weight of moisture to weight of bone dry material. The weight in pounds of product for any given feed rate and inlet and outlet moisture contents can then be expressed as $\frac{F\ (1\ +\ W_1)}{(1\ +\ W_o)}$. The outlet mass velocity of the gas at the top of the drier required to just float the particles was equal to the product of the free-falling velocity of the particle at the final moisture content, the cross-sectional area at the top of the drier, and the outlet density of the gas stream. In terms of the effective outlet mass velocity, the volume of gas flow will be $\frac{G'_2 A'}{\rho'}$. Division of the total product formed in a given time by the volume of gas that flows by it will give the hold-up:

$$H = \frac{F\ (1 + W_1)\ \rho'}{G'_2 A' (1\ +\ W_o)}$$

where: G'_2 = effective mass velocity at top of drier lb/hr/sq. ft.

ρ' = density of outlet gas stream, lb./ft.

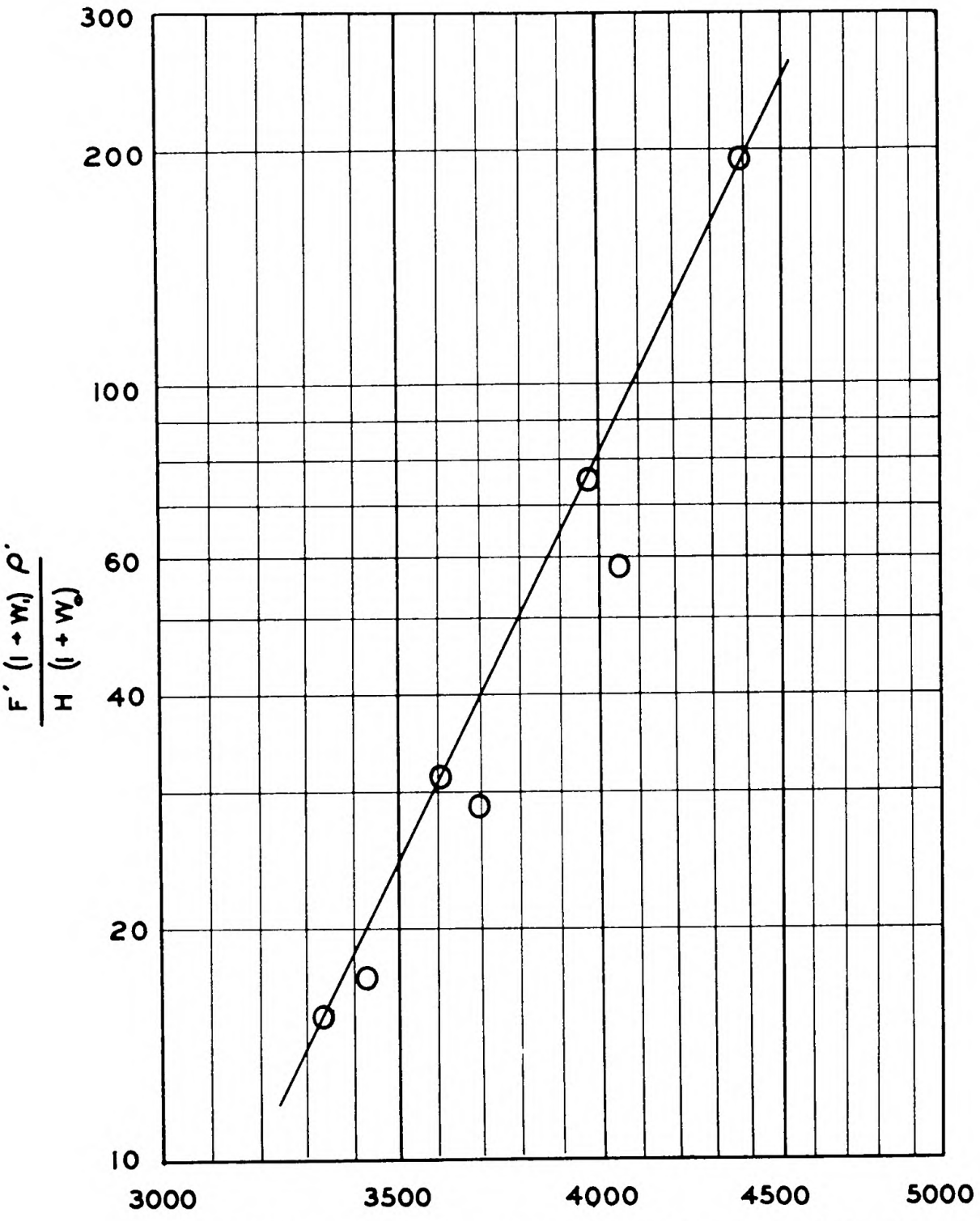


FIG. 3. $\frac{F'(1+W_1)\rho'}{H(1+W_0)}$ VS THE EFFECTIVE OUTLET MASS VELOCITY.

Letting $F' = F/A'$ and rearranging will give:

$$\frac{F' (1 + W_1) \rho'}{H (1 + W_o)} = G'_2$$

which applies only if the values of the effective mass velocity at the top of the drier are known accurately and the particles float in a stationary position in the rising gas stream. Since the particles do have an upward movement the two sides of the equation may be said to be proportional to each other but cannot be equal. Thus

$$\frac{F' (1 + W_1) \rho'}{H (1 + W_o)} \cong G'_2$$

The relationship between the effective outlet mass velocity and the expression $\frac{F' (1 + W_1) \rho'}{H (1 + W_o)}$ is given in Fig. 3.

The hold-up increases with an increase in feed rate if the remaining variables are held constant. Likewise, a decrease in the initial feed moisture content decreases the hold-up and a decrease in the product moisture content results in a greater hold-up if the remaining variables are constant. In terms of the effective mass velocity, feed rate, initial and final moisture contents, and the outlet gas stream density, the hold-up for the particle shape used under the conditions of this experimental work is given by the equation:

$$\frac{F' (1 + W_1) \rho'}{H (1 + W_o)} = \frac{G'^{9.088}_2}{6.47 \times 10^{30}}$$

The values of H, calculated from the above equation, are on the average within 9 per cent of the experimentally observed values for the dryer hold-up.

The values of k_G and h_G are dependent upon each other

and are expected to have corresponding changes in magnitude for any given change in operating conditions. During the experimental runs the mass velocity was a function of the temperature of the hot gas drawn in by the blower of essentially constant volume output. The mass velocity decreased as the temperature of the inlet gas stream was increased. Then as the mass velocity decreased, as shown in Fig. 2, the total surface area of transfer was increased, since the feed rate was approximately constant for all runs. The conditions are such that the experimental values of the transfer coefficients would increase with an increase in mass velocity. The experimental values of the mass and heat transfer coefficients versus the average mass velocity of the gas stream are shown in Figs. 4 and 5. The dependence of the transfer rates upon each other was borne out by the fact that the slopes of both curves were the same. This indicated that the two coefficients were related to each other so that if one were known the other could be readily calculated through the use of a common factor. This relationship is given by the equation:

$$h_G = 6.77\, k_G$$

which was based on the actual average mass velocity through the drier.

Since only the sensible heat contained in the gas that flowed past the particles caused evaporation of moisture, the values of the mass and heat transfer coefficients based on the actual mass velocity through the drier were not a true indication of the transfer rates. The mass velocity

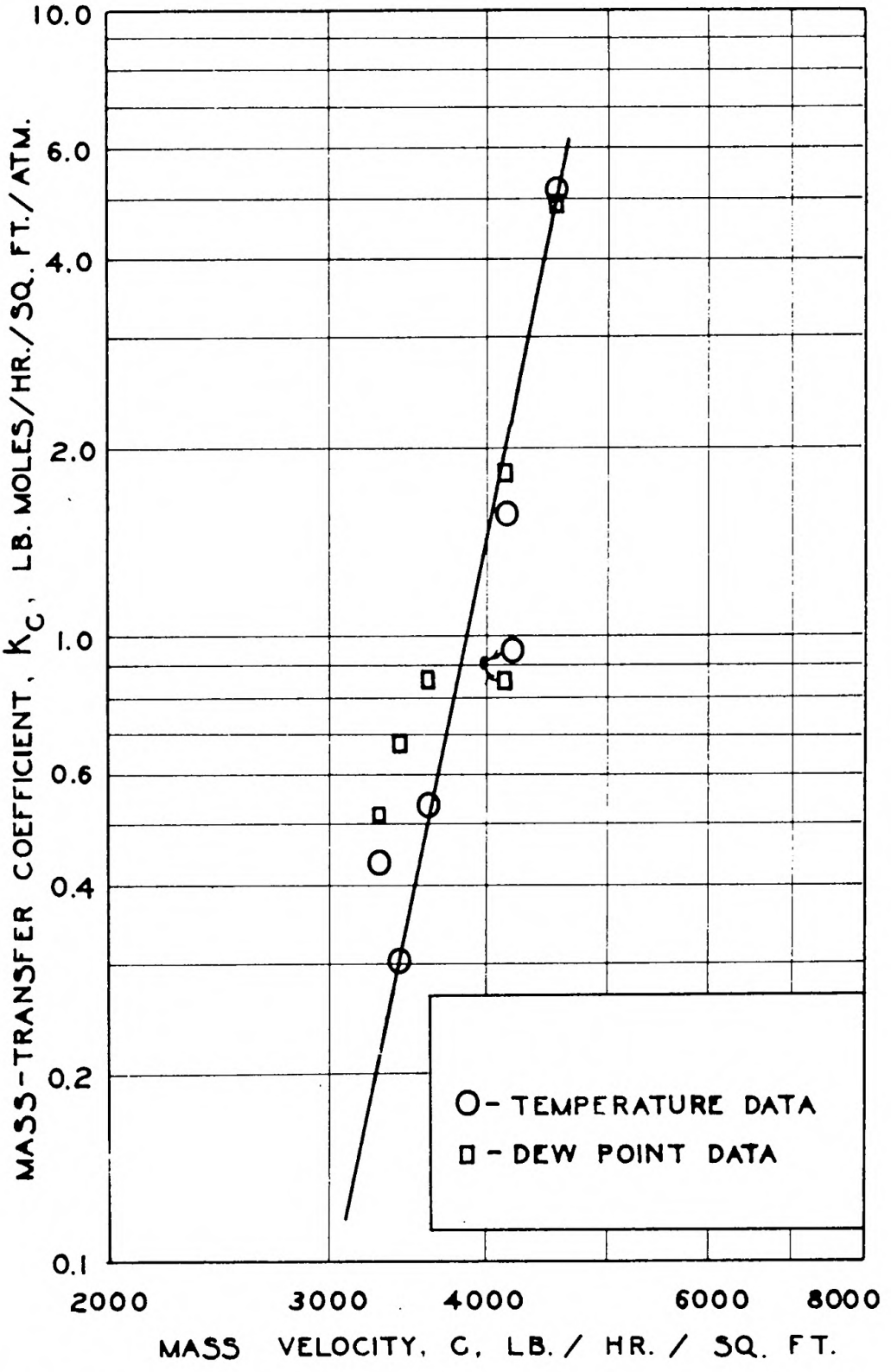


FIG. 4. MASS-TRANSFER COEFFICIENT FOR GAS FILM VS MASS VELOCITY THROUGH DRIER.

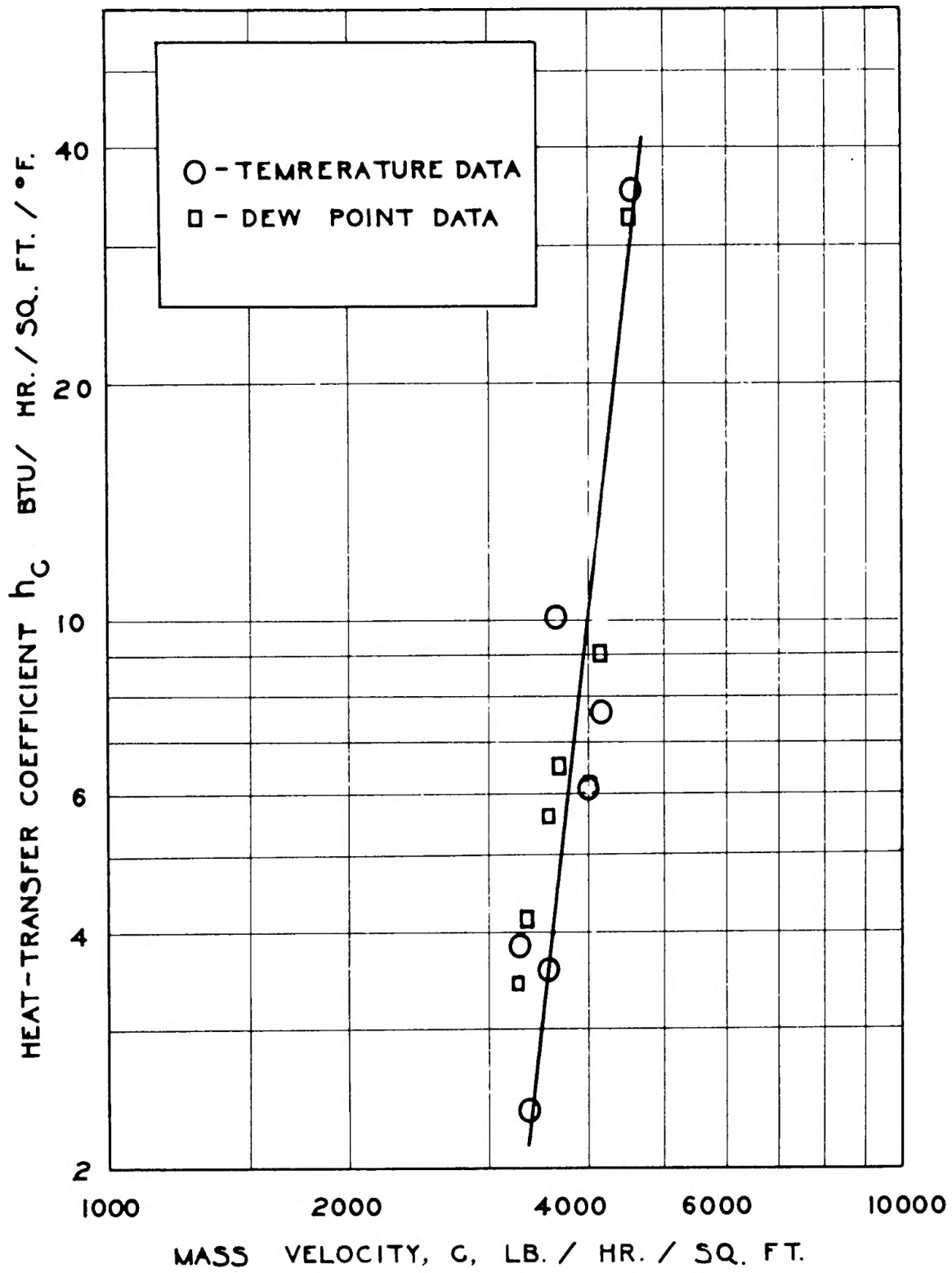


FIG. 5. HEAT-TRANSFER COEFFICIENT FOR GAS FILM VS MASS VELOCITY THROUGH DRIER.

available to sweep away the vapor envelope and supply the sensible heat for evaporation approached the values of the actual mass velocity at the lower flow rates. As the mass rate of flow increased the velocity of the particles through the drier increased from 122 feet per hour to 3221 feet per hour for the experimental range covered. Therefore, as the mass velocity increased an increased spread was obtained between the transfer coefficients based on the actual mass velocity through the drier and the transfer coefficients based upon the modified mass velocity through the drier. The difference between the values of k_G and h_G based on the actual and the modified mass velocities through the drier indicates the amount of error encountered by basing calculations for the transfer rates on the actual mass velocity. The slope of the curves shown in Figs. 6 and 7, where values of k'_G and h'_G are given versus the modified mass velocity through the drier, is 10.35 as compared to 8.5 obtained for the values of k_G and h_G shown in Figs. 4 and 5. These curves graphically illustrate the difference between the values of k_G and k'_G and h_G and h'_G. For any known value of k'_G or h'_G the corresponding value of h'_G or k'_G may be calculated from the following equation:

$$h'_G = 6.69\, k'_G$$

which was based upon the average modified mass velocity through the drier. The foregoing relationship applies only to the conditions investigated and the particular experimental drier used in this study.

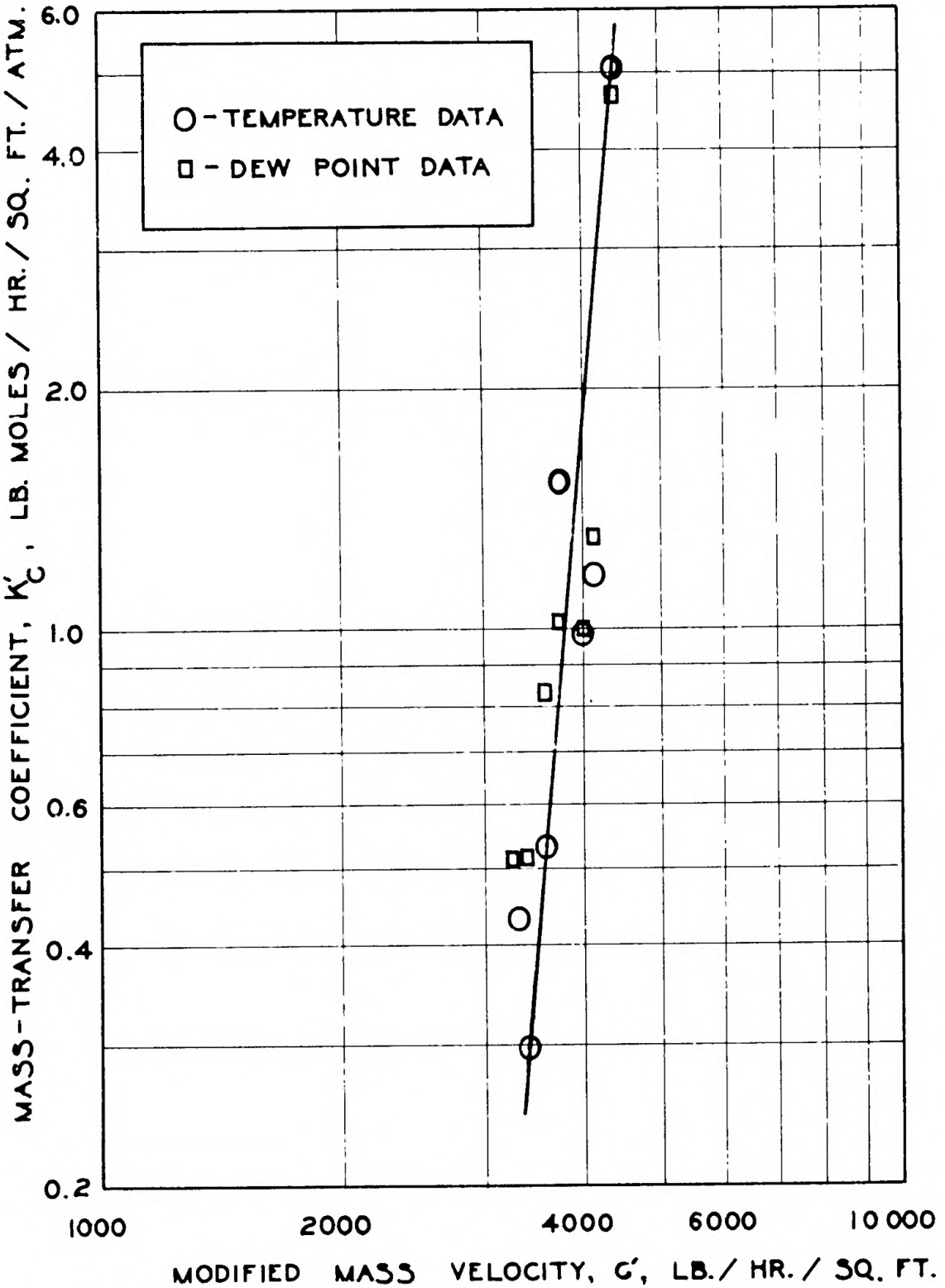


FIG. 6. MASS-TRANSFER COEFFICIENT FOR GAS FILM VS MODIFIED MASS VELOCITY THROUGH DRIER

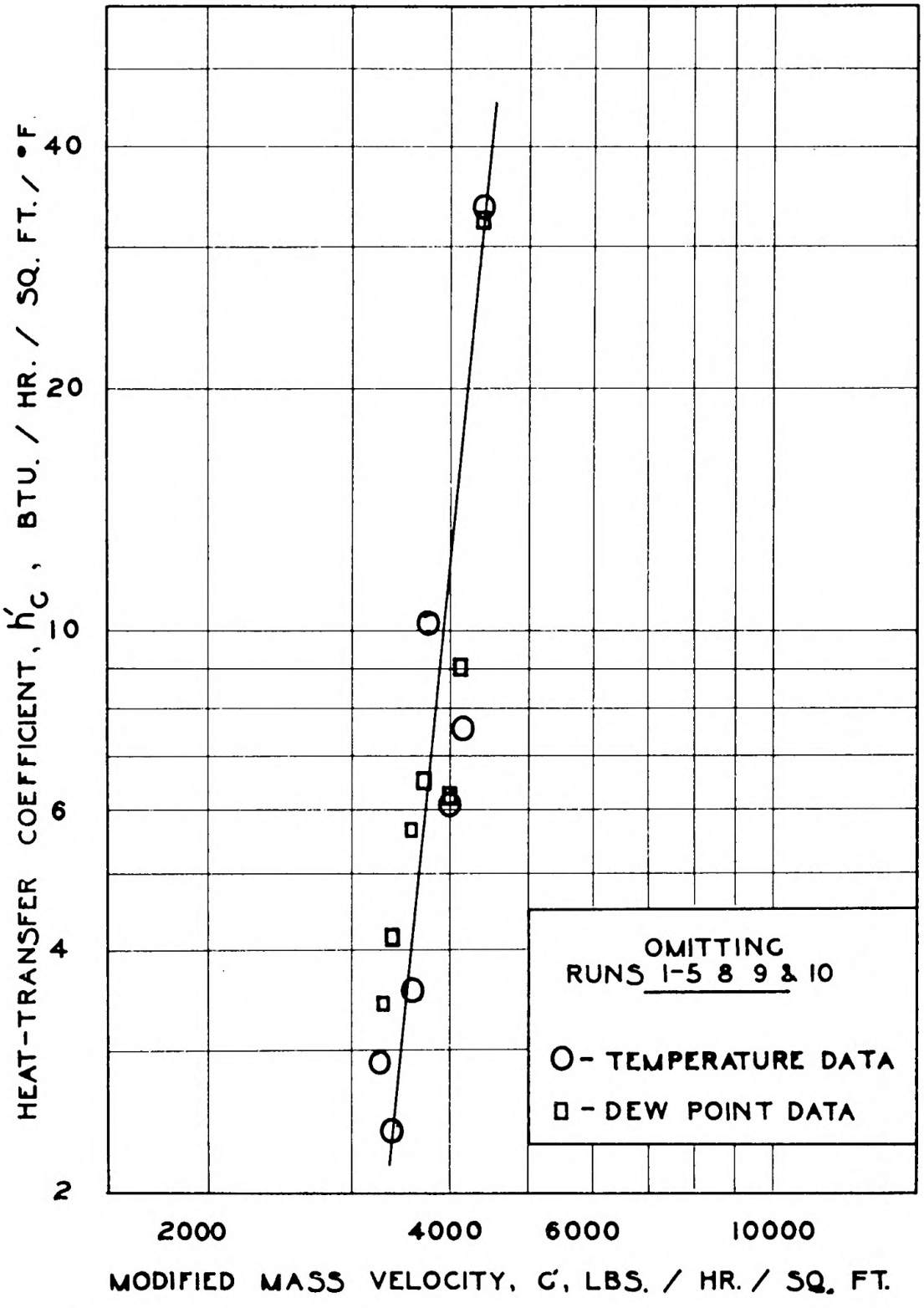


FIG. 7. HEAT-TRANSFER COEFFICIENT FOR GAS FILM VS MODIFIED MASS VELOCITY THROUGH DRIER.

The j factors for the mass and heat transfer through the gas film have the same dependence upon each other as did the transfer coefficients. This dependence is shown in Fig. 8. where the values of j'_d and j'_h are shown versus the modified Reynolds number display the same slope. Like the values of k_G and h_G the values of j_d and j_h would not be indicative of the true transfer values. In order to give a more reliable basis for estimation of the transfer rates the values of j'_d and j'_h, which are based on the modified mass velocity, are given versus the modified Reynolds number in Fig. 8. The values of j'_d and j'_h may be calculated from the following experimentally derived equations:

Equation	Average deviation from observed values
$j'_d = \left(\frac{D_p G'}{\mu}\right)^{5.654} \Big/ 2.37 \times 10^{23}$	35%
$j'_h = \left(\frac{D_p G'}{\mu}\right)^{5.654} \Big/ 1.82 \times 10^{23}$	36%

For particles over 200 microns in diameter Stokes law for free-falling velocities did not apply. The free-falling velocity for particles over 200 microns in diameter in arising turbulent gas stream are given by the following equation:

$$u = k_e \sqrt{d'' \left(\frac{\rho_s}{\rho} - 1\right)}$$

where: u = free-falling velocity of particle, ft/sec.

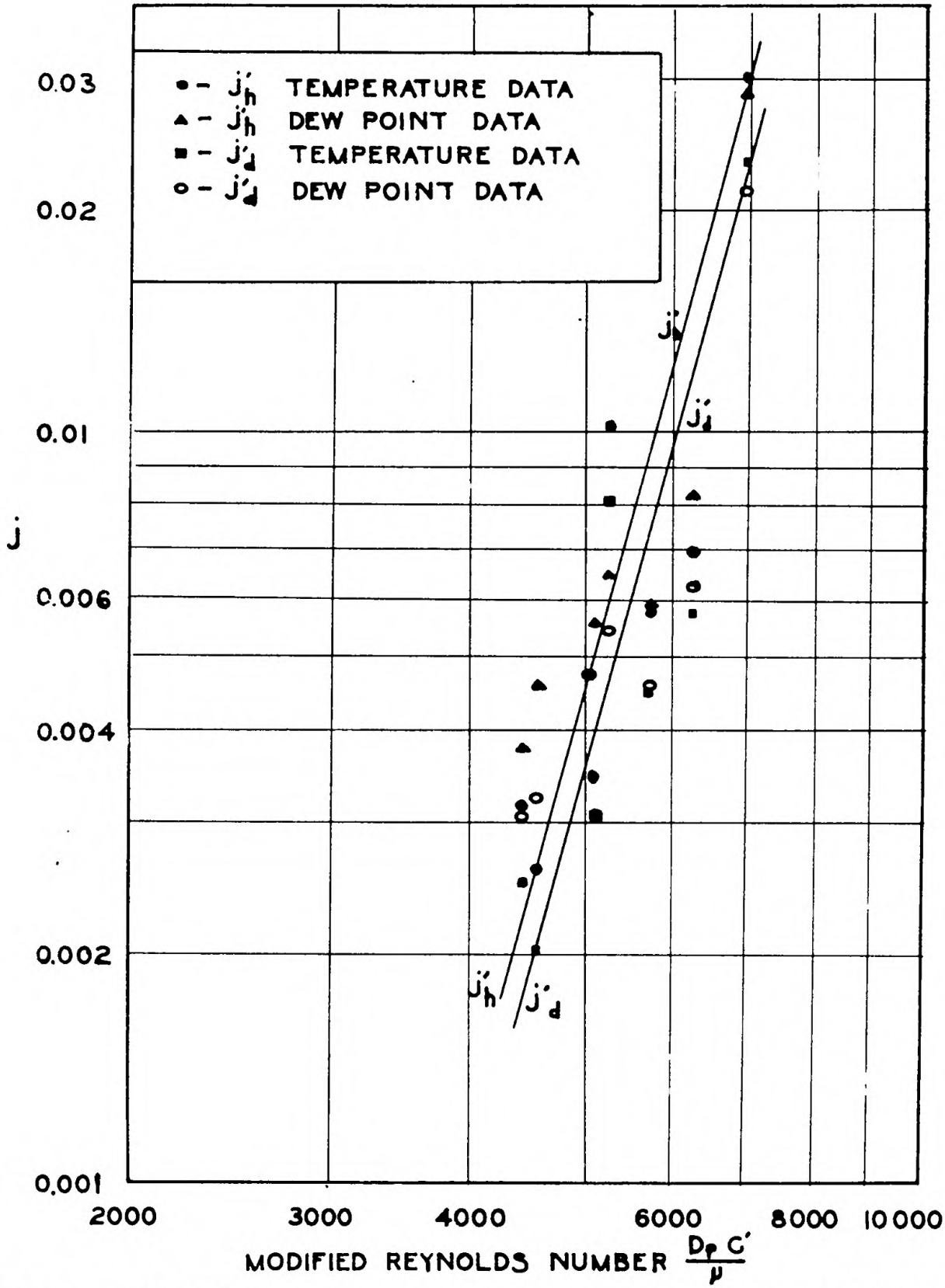


FIG. 8. J FACTORS FOR MASS-& HEAT-TRANSFER THROUGH GAS FILM VS THE MODIFIED REYNOLDS NUMBER.

d'' = particle diameter, inches

k_e = constant depending upon the particle shape

1.01 = disc or flaked material

The above equation will give the velocity to just float the particles in turbulent state. Since the particles do not remain stationary but do migrate through the drying section the recommended constant, $k_e = 1.01$, is not valid. For the range of conditions investigated the average experimental value of k_e is 0.503. This experimental value of k_e allows the calculation of the effective velocity past the particles.

$$u' = 0.503 \sqrt{d'' \left(\frac{\rho_s}{\rho} - 1\right)}$$

where: u' = effective velocity past the particles, ft./sec.

SUMMARY OF RESULTS

The sole purpose of this investigation was to obtain fundamental information for use in the design of floating bed flash driers. The experimental data for the drying during the constant-rate period of unwaxed, number 9, milk bottle caps were obtained on the veritcal drying section of a flash drier having a diameter of 16 inches for the first 7 feet and then increasing from 16 to 18 inches in diameter over the remaining 8 foot of length. The results for the experimental data are summarized as follows:

1. The hold-up in the drier was dependent upon the feed rate, inlet and outlet densities or moisture contents of the particles, and the mass velocity of the gas stream

at the top of the drying section. The hold-up in terms of the variables is given by the following equation:

$$\frac{F'}{H}\frac{(1 + W_1)\,\rho'}{(1 + W_o)} = \frac{G'^{9.088}_2}{6.47 \times 10^{30}}$$

where: F' = feed rate, lbs. /hr.- sq. ft. cross-sectional area at top of drier

ρ' = density of gas stream at top of drier, lb./ cu. ft.

H = hold-up, lbs. product / cu. ft. drier volume

G'_2 = mass velocity of gas stream at top of drier, lb. / hr - sq. ft.

W_o = moisture content of feed, lb. water / lb. bone dry material

W_1 = moisture content of product, bl. water / lb. bone dry material

2. The values of the mass-and heat-transfer coefficients vary over wide ranges with changes in mass velocity of the gas stream past the particles, the humidity of the drying gas, the gas stream temperature, particle size, and particle shape. Experimental values of the transfer rates were correlated in terms of the j **factors** which are expressions of the transfer rates in terms of the variables arranged in dimensionless groups. The j factors for mass-and heat-transfer are related in experimentally derived equations. They are:

$$j_d = \left(\frac{D_p G'}{\mu}\right)^{5.654} \Big/ \; 2.37 \times 10^{23}$$

$$j_h = \left(\frac{D_p G'}{\mu}\right)^{5.654} \Big/ \; 1.82 \times 10^{23}$$

where: D_p = effective particle diameter, ft.

G' = average modified mass velocity, lb./hr.-sq.ft.

μ = average viscosity of gas stream, lb./ft.-hr.

These foregoing relationships were derived from the experimental data on a flash drier that had a void fraction of 99 per cent or greater for all runs.

3. The effective velocity past the particles in the drier is given by the equation:

$$u' = 0.503\sqrt{d''\left(\frac{\rho_s}{\rho} - 1\right)}$$

where: u' = effective velocity past particle, ft/sec.

The foregoing relationships were derived from experimental data on a flash drier that had a void fraction of 99 percent or greater. No attempt was made to take into account the evaporation due to radient heat received by the particle or for the effect of the particles size and shape upon the transfer rates. Therefore, the use of these equations should be restricted to estimation of the transfer rates for flash drying, during the constant-rate period, for disc or flaked material having a diameter of about 1-5/16 inches at a maximum gas stream temperature of 500° F.

RECOMMENDATIONS FOR FUTURE WORK

The extremely delicate balance between the variables required for the design of floating bed flash dryers necessitates adequate knowledge of these variables and the extent to which each effects the drying operation.

Floating bed flash drying depends to a great extent upon the ability to float the dispersed particles in the rising turbulent gas stream. The mass velocity required to float the particles in the gas stream and give them a gradual upward velocity is a function of the free-falling velocity of the particles, therefore, further information relating particle size, shape, and density as a function of the free-falling velocity is necessary.

The mass-and heat-transfer rates are also dependent upon the particles size and shape as well as mass velocity past the particles, the temperature, and the humidity of the drying medium. With this in mind an interesting investigation would be to study the effect of temperature, mass velocity, and humidity of the gas stream on the rates of mass- and heat-transfer for various feed rates and particle size, shape, and density.

No information is available on the proper length for the vertical drying section of a floating bed flash drier. Information relating the variables to the rate of the particle movement through the drying section would facilitate the calculation of the optimum drying section length.

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APPENDIX

Table 1. Experimental and calculated data

Run	Length of run, secs.	Feed rate, lb./hr.	Gas consumed, cu. ft. / hr.	W_o	W_1	Temperatures, ° F. T_1	T_2	T_3	T_{DB}	T_{WB}	T_{WB}'	Temperature difference, ° F.	Water evaporated lb./hr. Y_1	Y_2	Y_3	Humidities, lb. H_2O/lb. B. D. air H_1	H_2	H_3	G	G_1	G_2	G'	c_p Inlet gas B.T.U./lb.-°F.	M_m	100 X viscosity, lb./ft.-hr. Ave. air stream	Mean gas film	$\frac{D_p G}{\mu}$	$\frac{D_p G'}{\mu}$
5	670	120.3	590	0.705	0.370	270	227	205	81	66	102	14	21.07	19.47	15.65	0.0161	0.0194	0.0192	4580	4580	4590	4400	0.244	28.69	5.06	3.74	7280	6960
6	633	116.6	873	0.425	0.207	350	326	290	81	69	116	13	17.60	17.65	13.55	0.0199	0.0231	0.0231	4040	4040	4050	3980	0.247	28.62	5.52	3.97	5880	5790
7	605	121.4	1410	0.545	0.071	490	438	386	78	65	127	32	37.23	66.14	12.82	0.0226	0.0306	0.0368	3410	3430	3440	3400	0.249	28.54	5.98	4.18	4610	4570
11	543	129.3	1380	0.472	0.056	440	377	320	86	71	122	23	28.17	44.26	18.95	0.0234	0.0291	0.0324	3610	3620	3630	3590	0.249	28.57	5.61	3.74	5180	5140
12	575	128.3	1410	1.120	0.639	520	454	354	88	73	130	57	64.09	76.68	56.30	0.0264	0.0408	0.0436	3280	3300	3330	3290	0.251	28.43	5.90	4.15	4500	4480
13	545	128.8	830	0.845	0.552	330	280	247	82	72	111	22	30.67	36.32	22.20	0.0208	0.0262	0.0268	4190	4200	4210	4140	0.246	28.61	5.27	3.85	6400	6300
14	607	127.5	1260	1.910	0.768	460	398	315	83	68	123	61	77.00	56.46	39.20	0.0219	0.0372	0.0318	3730	3760	3790	3740	0.248	28.49	5.69	4.07	5320	5280

Table 1. (cont.).

Run	D_v	Thermal conductivity 100 X k Ave. Air stream	Ave. film cond.	$\left(\frac{\mu}{\rho D_v}\right)_f^{2/3}$	$\left(\frac{c_p \mu}{k}\right)_f^{2/3}$	Based on temperature data: 100 X P_m, Atm.	k_G	k_G'	j_d'	h_G	h_G'	j_h'	Based on dew point data: 100 X P_m, Atm.	k_G	k_G'	j_d'	h_G	h_G'	j_h'	Heat loss from drier B.T.U./ hr. No feed	feed	ρ_s	$\rho_{s.}$	Hold up in drier: Total surface area, ft²	100 X lb. product per ft.³ drier.	k_e
5	1.22	1.83	1.38	0.703	0.951	4.10	5.24	5.05	0.023	35.20	33.80	0.030	4.12	4.83	4.67	0.021	32.50	32.50	0.029	7380	7200	58.80	47.20	5.44	2.19	0.51
6	1.37	2.04	1.49	0.705	0.946	7.00	0.91	0.90	0.005	6.16	6.08	0.006	7.10	0.90	0.90	0.005	6.19	6.19	0.006	12850	12650	48.80	40.30	15.28	5.42	0.58
7	1.46	2.23	1.58	0.738	0.944	12.30	0.30	0.30	0.002	2.37	2.37	0.003	9.43	0.69	0.51	0.003	4.17	4.17	0.005	18030	17850	53.20	35.80	56.20	17.80	0.50
11	1.44	2.13	1.54	0.720	0.893	8.17	0.54	0.53	0.003	3.57	3.54	0.003	8.10	0.85	0.84	0.005	5.62	5.62	0.006	14180	13200	53.90	38.80	35.70	11.60	0.53
12	1.53	2.26	1.58	0.706	0.940	9.95	0.43	0.43	0.002	2.86	2.88	0.003	9.90	0.51	0.51	0.003	3.43	3.43	0.004	16700	14400	73.40	56.80	83.33	24.40	0.41
13	1.31	1.94	1.44	0.698	0.944	5.26	1.60	1.17	0.006	7.61	7.53	0.007	5.30	1.77	1.31	0.006	9.02	9.02	0.008	10080	9320	63.20	53.00	20.32	7.77	0.53
14	1.45	2.15	1.54	0.711	0.943	7.79	1.57	1.53	0.008	10.02	10.02	0.010	8.50	0.96	1.02	0.005	6.48	6.48	0.006	15820	13050	98.00	59.70	34.16	10.65	0.40

See nomenclature for meaning of symbols in column headings.

SAMPLE CALCULATIONS

Run #12

Recorded Data

Operational Data

length of drying run	575	secs.
natural gas consumed	1410	cu. ft.
inlet temperature at blower	520	°F.
dry-bulb temperature	88	°F.
wet-bulb temperature	73	°F.
static pressure upstream orifice	3.0	in. water
" " down stream orifice	2.2	in. water
" " feed level	2.3	in. water
" " top drying section	2.1	in. water
diameter of orifice	7	in.
weight of feed	20.50	lbs.
number of caps hold-up in drier	4705	
pressure drop across orifice	1.17	in. water

Analysis Data

Analyzer No.	Sample level ft. from top	Temp., Analyzer ° F.	H in in. Hg	Analyzer M. V.
1	7	113	19.44	4.25
2	4	107	18.06	3.95
3	1	111	21.69	4.37
4	9	108	19.45	3.68
5	16	110	19.50	3.85
6	10	104	21.25	4.24
7	14	103	17.94	3.71
8	12	105	20.81	3.76

Drying Section Data

Recorder Pt. No.	Level ft. from top	No Feed Scale Rdg.	No Feed Temp. ° F.	Feed Scale Rdg.	Feed Temp. ° F.
1	1	534	411	385	361
2	2	546	414	394	365
3	4	561	419	398	366
4	5	564	420	409	369
5	7	556	418	425	374
6	9	592	429	454	384
7	10	604	433	470	389

Recorder Pt. No.	Level ft. from top	No Feed Scale Rdg.	No Feed Temp. ° F.	Feed Scale Rdg.	Feed Temp. ° F.
8	11	613	436	470	389
9	12	615	437	480	393
10	13	632	442	475	391
11	14	642	446	479	392
12	15	653	454	457	386
13	16	653	454	653	454
14	top drier wall	104	260	7	237
15	mid. drier wall	109	270	25	243
16	bottom drier wall	78	252	44	249

Calculated Data

Moisture content of inlet gas stream

Analyzer temp. = 3.85 M. V. = 163.4 ° F.

From plot of analyzer temperature us the dew point temperature an analyzer temperature of 163.4 ° F. equals a dew point of 80.0 ° F.

moisture content of analyzer mix = 0.0222 $\frac{\text{lb. water}}{\text{lb. B. D. air}}$

moisture content of room air = 0.0139 $\frac{\text{lb. water}}{\text{lb. B. D. air}}$

volume fraction sample = 19.50/29.20 = 0.669

moisture content of sample

$$\frac{0.0222 + (1 - 0.669)(0.0139)}{0.669} = 0.0264 \frac{\text{lb. water}}{\text{lb. B.D. air}}$$

Heat loss from drier wall, no feed

room temperature	88 ° F.
average wall temperature	263 ° F.
(ΔT_s)	175 ° F.
D_o'	16 inches

$$h_c = 0.50 \ (\Delta T_s / D_o') \ 0.25 =$$

$$= 0.50 \ (\ 175/16) \ 0.25 = 0.909$$

From fig. 12 page 983 of Perry's Handbook the value of h_r at a wall temperature of 263° F. is 0.58 for an emisivity of 1. The emisivity for the drier wall is 0.276 as taken from table 1 page 1005 of the Chemical Engineers Handbook.

$$Q = (h_c + h_r) A (\Delta T) =$$

$$= (0.909 + 0.58) \ 64.02(175) = 16700 \text{ B.T.U./ hr.}$$

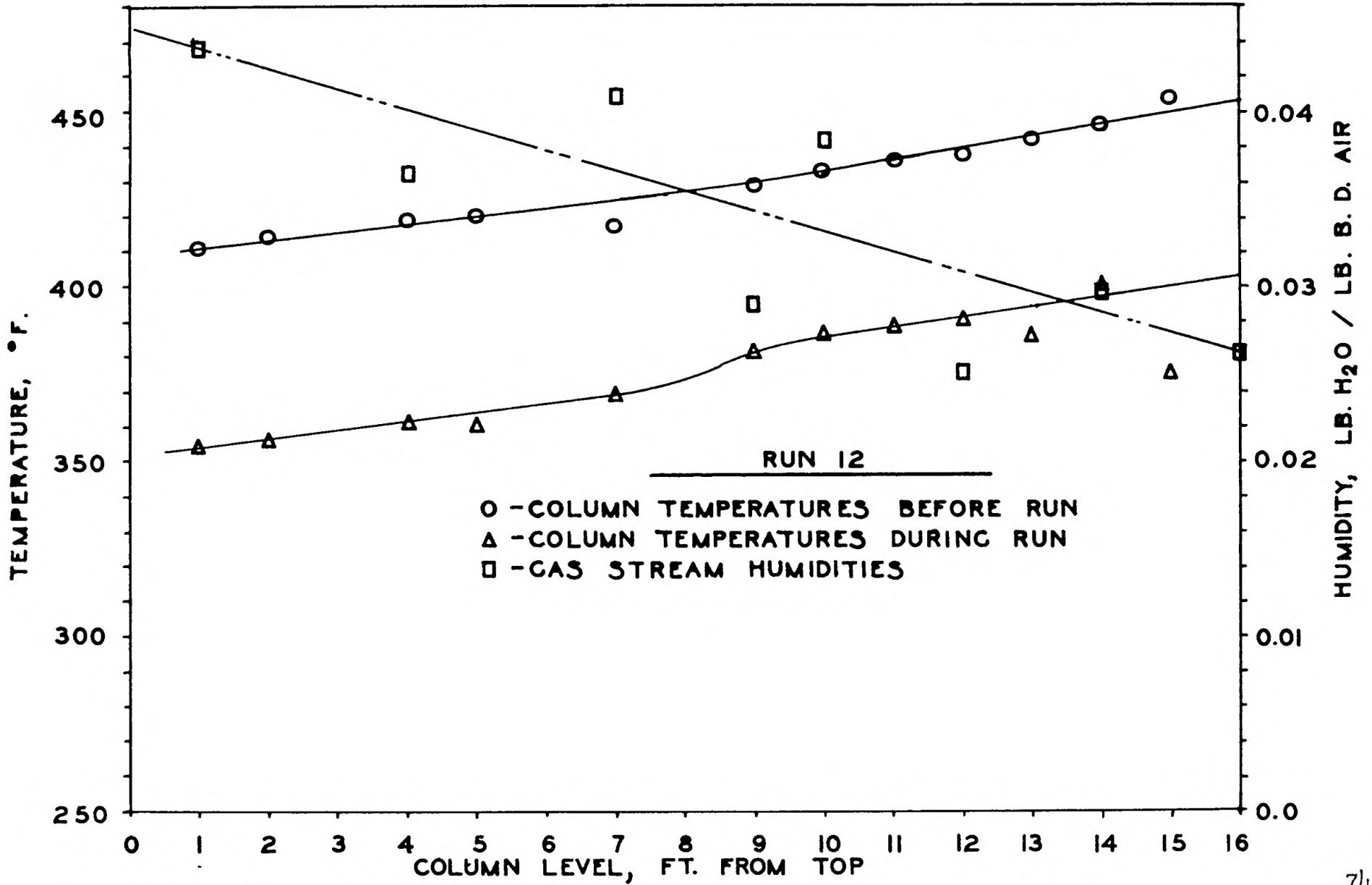

FIG. 9. GRAPHICAL PRESENTATION OF DATA ON RUN 12.

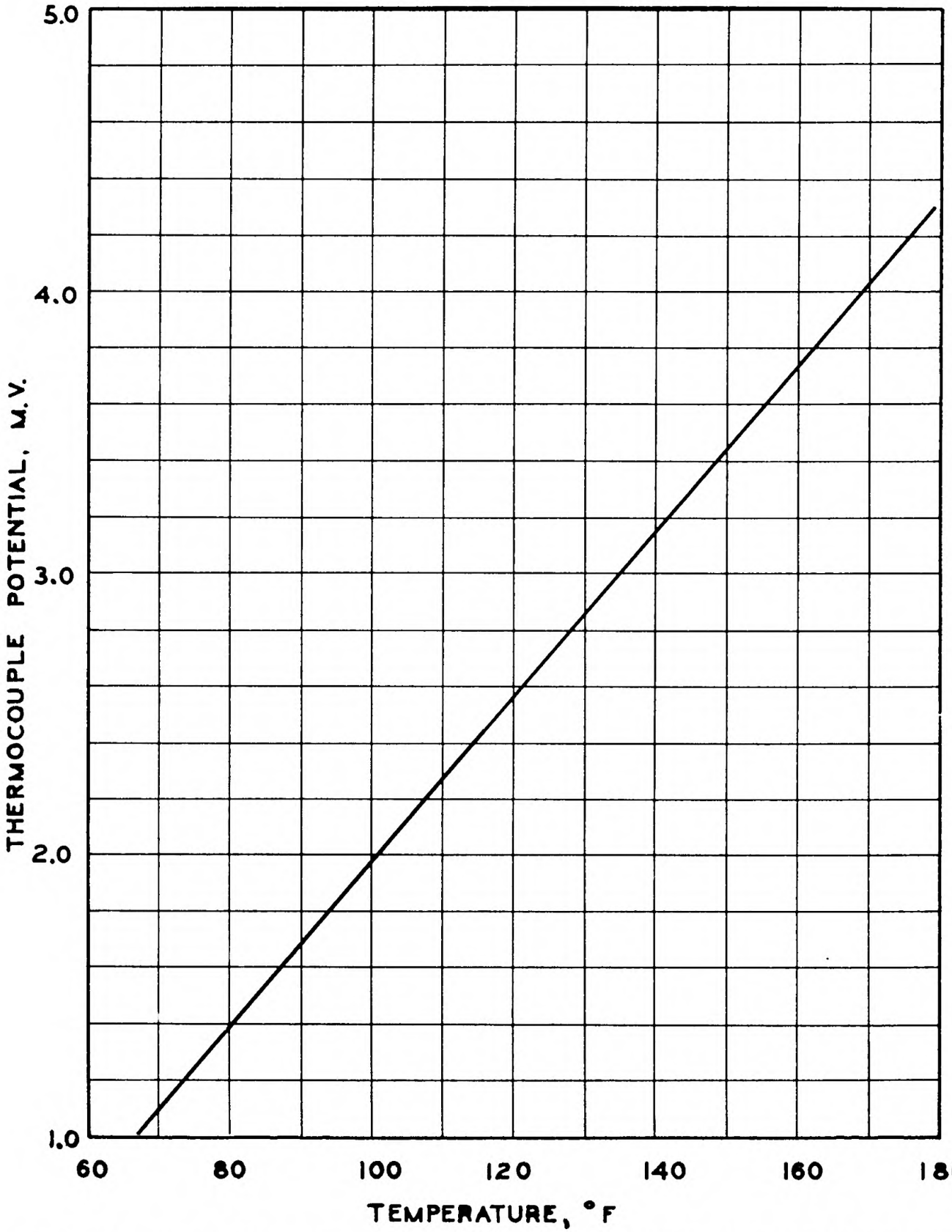

FIG. 10. CALIBRATION CURVE IRON-CONSTANTAN THERMOCOUPLES

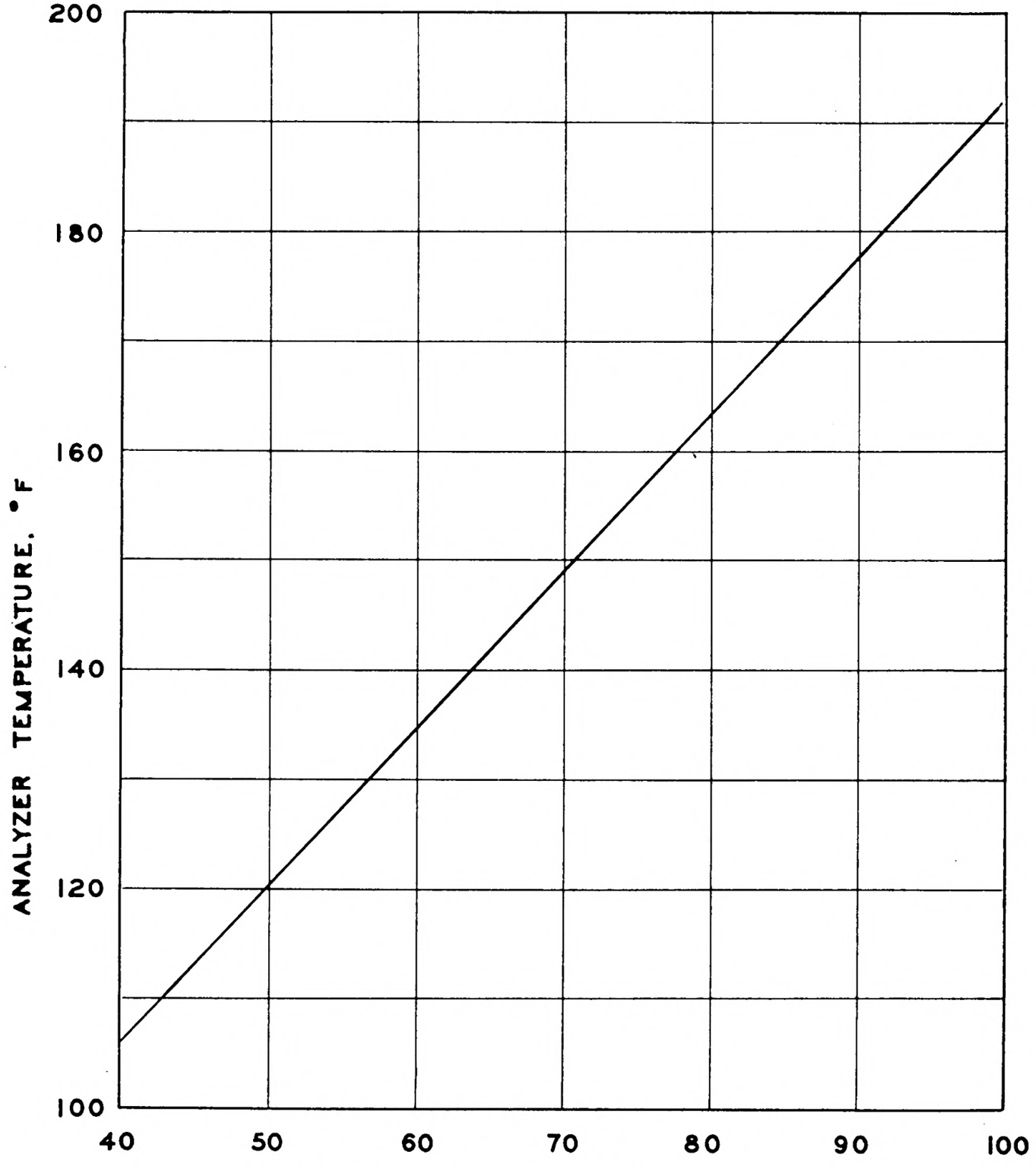

FIG. 11. ANALYZER CALIBRATION CURVE.

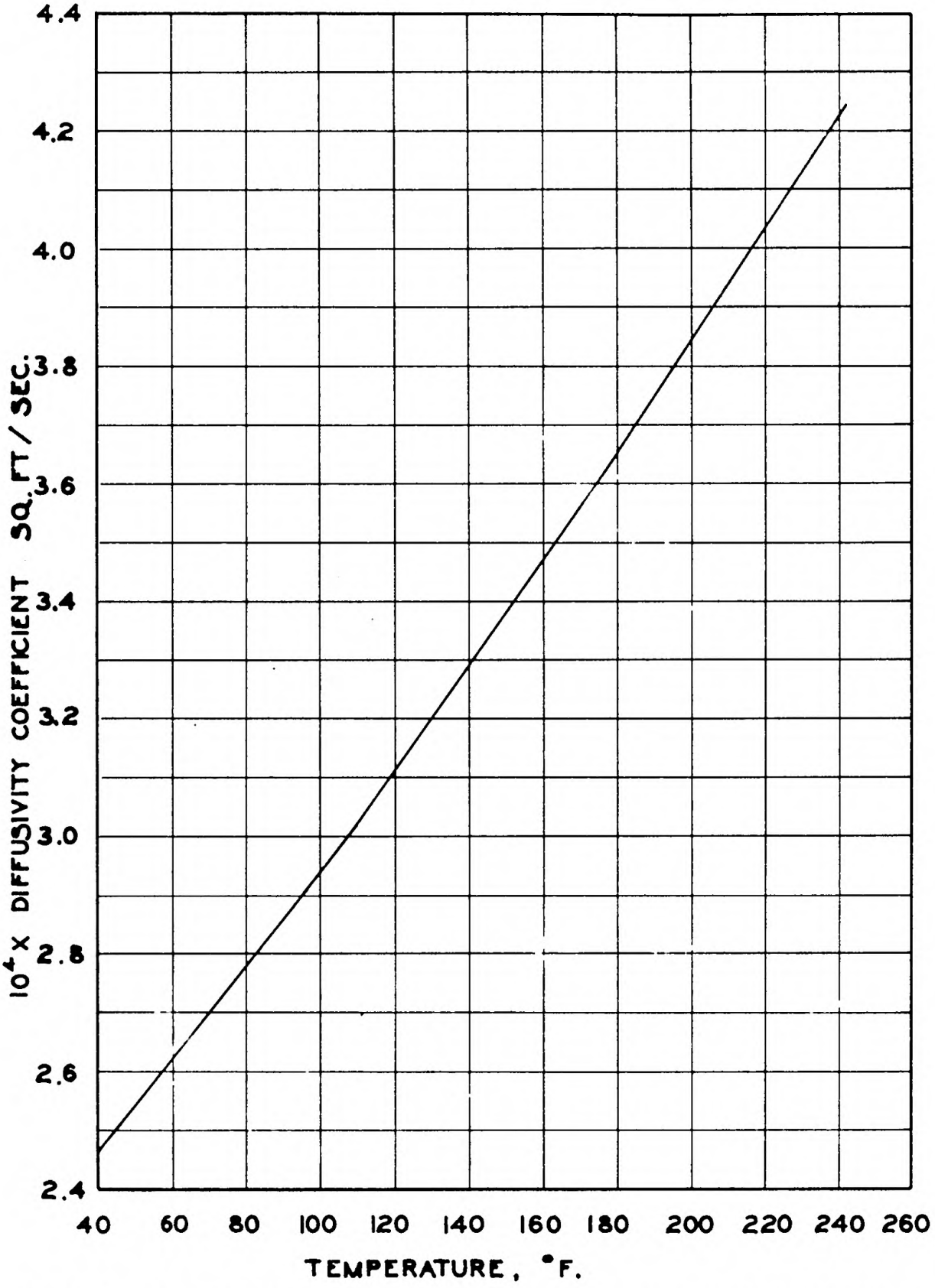

FIG. 12. DIFFUSIVITY COEFFICIENTS FOR AIR-WATER VAPOR MIXTURES AT 1 ATMOSPHERE.

Heat loss from drier wall, with feed

average wall temperature 243 ° F.
(T_s) 155 ° F.

$h_c = 0.50 (155/16) 0.25 = 0.882$

$h_r = 2.05 (0.276) = 0.566$

$Q = (0.882 + 0.566)64.02(155) = 14400$ B.T.U./hr.

Inlet mass velocity through drier, G

$$G = \frac{3600\, C_o\, Y\, S_2}{A} \sqrt{\frac{2 g_o \rho^2 \Delta H}{(1-\beta^4)}}$$

$$= \frac{3600(0.78)(0.99)(0.267)}{1.395} \sqrt{\frac{2(32.17)(0.0402)^2\ 151.2}{0.413}}$$

$= 3280$ lb./ hr. -sq. ft.

Outlet mass velocity through drier, G_2

mean C_p at 400 ° F. $= \frac{0.245 + (0.463)(0.0264)}{1.0264} =$

$= 0.255$ B.T.U./lb. - ° F.

water vaporized, $Y_1 = \frac{3280\ (1.395)(0.255)57}{1019} = 64.09$ lb./hr.

$$G_2 = \frac{3280(1.395) + 64.09}{1.395} = 3326 \text{ lb./hr.-sq. ft.}$$

Average mass velocity through drier, G_1

$$G_1 = \frac{3280 + 3326}{2} = 3300 \text{ lb./hr.-sq. ft}$$

Modified mass velocity, G'

wt. per cap. 0.607 grams B. D.
surface area per cap = 0.02032 $ft.^2$

cap hold-up in drier = 4705/24.47 = 192 caps/ft^3 drier volume

$192(0.607)/453.6 = \frac{0.257 \text{ lb. B. D. mat'l}}{\text{cu. ft. tower volume}}$

128.3/ (1 + 1.92) = 43.95 lb B. D. mat'l/ hr.

velocity of caps through drier $= \frac{43.95}{(0.257)(1.395)} = 122 \frac{\text{ft.}}{\text{hr.}}$

$H_2 = 1.0264 + 64.09/4455 = 1.0408 \frac{\text{lb. water}}{\text{lb. B. D. air}}$

average density of gas stream

$$= \frac{0.0529 + 0.0402}{2} = 0.0479 \text{ lb./ cu. ft.}$$

$$G' = \left[\frac{3300}{0.0479} - 122\right] 0.0479 = 3292 \text{ lb./ hr. - sq. ft.}$$

Modified outlet mass velocity, G_2'

$$G_2' = \left[\frac{3326}{0.0529} - 122\right] 0.0529 = 3320 \text{ lb./hr.-sq. ft.}$$

Mass-transfer coefficient based on temperature data, k_G

$$r_{A_a} = \frac{64.09}{18\ (192)(0.02032)(21.4)} = 0.043 \frac{\text{lb moles}}{\text{hr. -ft}^2}.$$

$k_G = r_{A_a} / (\Delta P)_m$

$$(\Delta P)_m = \frac{\Delta P_1 - \Delta P_2}{\ln(\Delta P_1/\Delta P_2)}$$

$$\text{p inlet} = \frac{29(0.0264)}{18 + 29\ (0.0264)} = 0.0408 \text{ atm.}$$

$$\text{p outlet} = \frac{29(0.0408)}{18 + 29(0.0408)} = 0.0617 \text{ atm.}$$

p wet-bulb temp. = 0.1510 atm.

$\Delta P_1 = (0.1510 - 0.0408) = 0.1102$ atm.

$\Delta P_2 = (0.1510 - 0.0617) = 0.0893$ atm.

$(\Delta P)_m = (0.1102 - 0.0893)/ \ln(0.1102/ 0.0893)$

$= 0.0995$ atm.

$k_G = 0.043/ 0.0995 = 0.432$ lb. moles/hr.-sq.ft. - atm.

Mass-transfer coefficient based on dew point data, k_G

$H_3 = 0.0436$ lb. water/ lb. B. D. air

water vaporized $= (0.0436 - 0.0264)\ 4455 = 76.68 \frac{\text{lb.}}{\text{hr.}}$

$(\Delta P)_m = (0.1102 - 0.0853)/\ln(0.1102/0.0853)$

$= 0.099$ atm.

$k_G = 76.68/\ 18(83.33)(\ 0.099) = 0.51$ lb. moles / hr.-sq. ft. -atm.

Mass-transfer coefficient from temperature data based on the modified mass velocity through the drier, k_G^1

$$Y_1 = \frac{3292\ \ (1.395)57(0.251)}{1019} = 64.301\text{b./hr.}$$

$k_G^1 = 63.4/18(83.33)(0.099) = 0.43$ lb. moles / hr.-sq. ft.-atm.

Mass-transfer coeffieient from dew point data based on the modified mass velocity through the drier, k_G^1

$k_C = 76.68/\ 18(83.33)(0.099) = 0.051$ lb. moles / hr.-sq. ft.-atm.

Mean molecular weight of gas stream, M_m

basis: 1 lb. B. D. air

moles air = 0.0345 lb moles

average moles water = $\frac{0.0408 + 0.0264}{2(18)} = 0.0019$ lb. mole

mole fraction B. D. air = 0.984

mole fraction water vapor = (1 - 0.984) = 0.052 lb. moles

ΔMW air = 29(0.984) = 27.49
ΔMW water = 18(0.052) = 0.94

$M_m = 28.43$

Film pressure factor, P

$P_{B_2} = (\pi - p_{A_2}) = (1.005 - 0.0408) = 0.964$ atm.

$P_{B_1} = (\pi - p_{A_1}) = (1.005 - 0.0617) = 0.943$ atm.

$P_f = (0.943 - 0.964)/\ln(0.943/0.964) = 0.963$ atm.

Effective particle diameter, D_p

$D_p = \sqrt{A_p\ /3.1416} = \sqrt{0.02032/\ 3.1416} = 0.0804$ ft.

Mean gas stream viscosity,

mole fraction air, inlet = 0.958

mole fraction air, outlet = 0.937

mole fraction water vapor, inlet = 0.042

mole froction water vapor, outlet = 0.063

T_c air = 238.2° R. T_c water vapor = 1165 ° R.

P_c air = 37.2 atm. P_c water vapor = 217.7atm.

T_c' = (N air T_c air) + (N water vapor T_c water vapor)

T_c' = (228 + 48.9) = 277 ° R. inlet

T_c' = (223 + 73.4) = 296 ° R. outlet

P_c' = (N air P_c air) = (N water vapor P_c water vapor)

P_c' = (35.6 + 9.14) = 44.7 atm.

P_c' = (34.8 + 13.71) = 48.5 atm.

inlet gas stream temperature = 914 ° R.

outlet gas stream temperature = 814 ° R.

T_r' inlet = T inlet/ T_c' inlet = 914/277 = 3.30

T_r' outlet = 814/296 = 2.75

P_r' inlet = P/P_c' = 1.005/ 44.7 = 0.0224

P_r' outlet = 1.005/ 48.5 = 0.0207

From the generalized reduced viscosities, the reduced viscosity for the inlet and outlet conditions were obtained for the respective values of T_r' and P_r'.

μ_r inlet = 1.24 μ_r outlet = 1.10

μ_c' = (N air μ_c air) + (N water vapor μ_c water vapor)

μ_c' = (0.985)(193) + (0.042)(495) = 206 micropoise, inlet

μ_c' = (0.937)(193) + (0.063)(495) = 212 micropoise, outlet

μ inlet = $\mu_r \mu_c'$ = 1.24(2.06) = 2.55 micropoise

μ outlet = 1.10(2.12) = 2.33 micropoise

Mean gas stream viscosity = (2.55 + 2.33)/2 = 2.44 micropoise

2.44 micropoise = 0.059 lb./ft.-hr.

Mean viscosity of gas film

viscosity of water vapor at wet-bulb temperature

$$\mu = 0.0278 \text{ lb./ft.-hr.}$$

mean gas film viscosity = $\frac{(0.059 - 0.0278)}{\ln \frac{0.059}{0.0278}} = 0.0415 \frac{\text{lb.}}{\text{ft.-hr.}}$

Mean temperature of gas film

mean temperature of gas stream = (454 + 354) / 2 = 404° F.

wet-bulb temperature of entering gas stream = 130 ° R.

mean film temperature =

(404 - 130)/ ln (404/130) = 242 ° F.

Mean diffusivity of gas film

From plot of diffusivity coefficients vs temperature for air-water vapor mixtures

D_v at 242 ° R. = 0.000425 sq. ft./sec.

D_v = 0.000425(3600) = 1.53 sq. ft./ hr.

Mean density of gas film

density of water vapor at wet-bulb temperature
= 0.0418 lb./ cu. ft.

density of inlet gas stream = 0.0432 lb./ cu. ft

density of outlet gas stream = 0.0529 lb./ cu. ft

average density of gas stream = $\frac{(0.0432 + 0.0529)}{2}$ =

= 0.0458 lb./cu. ft.

j_d' for mass-transfer through the gas film based on temperature data

$$j_d' = \left(\frac{0.43(0.963)28.42}{3290}\right)\left(\frac{0.0415}{0.0458(1.53)}\right)_s^{2/3} = 0.0025$$

j'_d for mass-transfer through the gas film vased on dew point data

$$j'_d = \left(\frac{0.51\ (0.963)\ 28.43}{3290}\right) 0.706 = 0.00304$$

Heat-transfer coefficients

Based on temperature data

$$h_G = Q/(A\,\Delta T) = 65305/\ [(83.33)274] = 2.86\ \frac{\text{B.T.U.}}{\text{hr.-ft.}^2\text{-}^\circ\text{F}}$$

$$h'_G = 65600/\ [(83.33)274] = 2.88\ \frac{\text{B.T.U.}}{\text{hr.- ft.}^2\text{ - }^\circ\text{F.}}$$

Based on dew point data

$$h_G = 78100/(83.33)\ 274 = 3.43.\ \frac{\text{B.T.U.}}{\text{hr.- ft.}^2\text{- }^\circ\text{F.}}$$

$$h'_G = 78500/\ (83.33)\ 274 = 3.44\ \frac{\text{B.T.U.}}{\text{hr.-ft.}^2\text{ - }^\circ\text{F.}}$$

Mean thermal conductivity of gas film

k = (N air k air + N water vapor k water vapor)

k inlet = (0.0230 + 0.0009) = 0.0239 $\frac{\text{B.T.U. ft.}}{\text{hr. - ft.}^2\text{ - }^\circ\text{F.}}$

k outlet = (0.0202 + 0.0012) = 0.0214 $\frac{\text{B.T.U. ft.}}{\text{hr. - ft.}^2\text{ - }^\circ\text{F.}}$

k water vapor at wet-bulb temperature
= 0.0108 $\frac{\text{B.T.U. ft.}}{\text{hr. - ft.}^2\text{ - }^\circ\text{F.}}$

$$k = (0.0214 - 0.0108)/\ln (0.0214/0.0108) = 0.0155\ \frac{\text{B.T.U. ft.}}{\text{hr. - ft.}^2\text{ - }^\circ\text{F.}}$$

Mean C_p gas film

C_p average gas stream = 0.255 B.T.U./ lb.- $^\circ$ F/

C_p water vapor at wet-bulb temperature = 0.462 $\frac{\text{B.T.U.}}{\text{lb. - }^\circ\text{F.}}$

$$C_p \text{ mean} = \frac{(0.462 - 0.255)}{\ln (0.462/0.255)} = \frac{0.346\ \text{B.T.U.}}{\text{lb. - }^\circ\text{F.}}$$

j'_h for heat-transfer based on temperature data

$$j'_h = \left(\frac{h_G}{C_p\ G'}\right) \left(\frac{C_p\ \mu}{k}\right)^{2/3}_f$$

$$= \left(\frac{2.88}{(0.255)3292}\right)\left(\frac{(0.346)\ 0.0415}{0.0155}\right)^{2/3}_{s} = 0.0032$$

j'_h for heat-transfer based on dew point data

$$j'_h = \left(\frac{3.43}{(0.255)\ 3292}\right) \quad (0.940) = 0.0038$$

Evaluation of k_e

$$u' = k_e\sqrt{d''\left(\frac{\rho_s}{\rho} - 1\right)}$$

ρ_s = 73.4 lb./ cu. ft.

ρ = 0.0432 lb/ cu. ft.

u' = 70790/ 3600 = 19.65 ft. / sec.

$$k_e = \frac{19.65}{\sqrt{1.313\left(\frac{73.4}{0.0432} - 1\right)}} = 0.41$$

NOMENCLATURE

A - Transfer area, sq. ft

A' - Cross-sectional area of drier, sq. ft.

A_p - Surface area per particle, sq. ft.

B. D.- Bone dry.

C_o - Orifice coefficient, dimensionless.

C_p - Specific heat, B.T.U./lb. - ^{o} F.

D_1 - Diameter of duct, ft.

D_2 - Diameter of orifice opening, ft.

D_p - Effective diameter of particle, ft.

D_v - Average diffusivity coefficient, sq. ft./hr.

d'' - Diameter of particle, inches

F - Feed rate, lbs./ hr.

F' - Feed rate, lbs./ hr. - sq. ft. drier cross-section.

G - Mass velocity of gas entering drier, lb./hr.-sq. ft.

G_1 - Average mass velocity of gas through drier, lb. / hr. - sq. ft.

G_2 - Outlet mass velocity through drier, lb. / hr.-sq. ft

G' - Average effective mass velocity through drier, lb. / hr.-sq. ft.

G_2' - Effective outlet mass velocity through drier, lb. / hr. - sq. ft.

g_c - Dimensional constant, $32.17 \frac{(\text{lb. mass})(\text{ft.})}{(\text{lb. force})(\text{sec.}^2)}$

H - Hold-up, lbs. product/ cu. ft. drier volume.

H_1 - Humidity of entering gas stream, $\frac{\text{lb. water}}{\text{lb. B. D. air}}$

H_2 - Hunidity of outlet gas stream from temperature data $\frac{\text{lb. water}}{\text{lb. B. D. air}}$

H_3 --Humidity of outlet gas stream from dew point data, lb. water/ lb. B. D. air.

H - Orifice differential, ft. of fluid flowing

h_c - Convection heat transfer coefficient, $\frac{\text{B. T. U.}}{\text{hr. - sq. ft. - }^{o}\text{F.}}$

h_G - Heat transfer coefficient based on the actual mass velocity through the drier, $\frac{\text{B. T. U.}}{\text{hr. - sq. ft. - }^{o}\text{F.}}$

h'_G - Heat transfer coefficient based on dew point data, $\frac{\text{B. T. U.}}{\text{hr. - sq. ft - }^{o}\text{F.}}$

h_r - Radiation heat transfer coefficient, $\frac{\text{B.T.U}}{\text{hr.- sq. ft. - }^{o}\text{F.}}$

j_d - Mas transfer factor based on the actual mass velocity through the drier, dimensionless.

j'_d - Mass transfer factor based on the modified mass velocity through the drier, dimensionless.

j_h - Heat transfer factor based on the actual mass velocity through the drier, dimensionless.

j'_h - Heat transfer factor based on the modified mass velocity through the drier, dimensionless.

k - Thermal conductivity, $\frac{\text{B.T.U. - ft.}}{\text{hr. - sq. ft. - }^{o}\text{F.}}$

k_e - Constant depending upon the shape of the particle.

k_f - Thermal conductivity of gas film, $\frac{\text{B.T.U. - ft.}}{\text{hr. - sq. ft. - }^{o}\text{F.}}$

k_G - Mass transfer coefficient of gas film based upon the actual mass velocity through the drier, lb. moles/ hr. - sq. ft. - atm.

k'_G - Mass transfer coefficient of gas film based upon the modified mass velocity through the drier, lb. moles / hr. - sq. ft. - atm.

L - Length of drying section, ft.

M_m - Mean molecular weight of gas stream, lbs. / mole.

P_f - Film pressure factor, atm.

P_{A_1} - Partial pressure of water vapor in entering gas stream, atm.

P_{A_2} - Partial pressure of water vapor in exit gas stream, atm.

$(\Delta P)_m$ - Mean partial pressure difference for water vapor, atm.

Q - Heat to vaporize water, B.T.U. / hr.

Q_1 - Heat loss from drier wall before feed, B.T.U /hr.

Q_2 - Heat loss from drier wall during run, B.T.U./ hr.

S_2 - Cross-sectional area of orifice opening, sq. ft.

T_1 - Temperature of gas stream at blower, ^{o} F.

T_2 - Temperature of gas stream entering drier, ^{o} F.

T_3 - Temperature of gas stream leaving drier, ^{o} F.

T_{DB} - Dry-bulb temperature of room air, ^{o} F.

T_{WB} - Wet-bulb temperature of room air, ^{o} F.

$T_{WB'}$ - Wet-bulb temperature of entering gas stream, ^{o} F.

$(\Delta T)_m$ - Mean temperature difference between inlet and outlet conditions, ^{o} F.

u - Free-falling velocity of particle, ft. / sec.

u' - Effective mass velocity past particle, ft. / sec.

V - Velocity of gas stream, ft. / hr.

W_o - Initial moisture content of particles, $\frac{\text{lb. water}}{\text{lb. B. D. mat'l}}$

W_1 - Final moisture content of particle, $\frac{\text{lb. water}}{\text{lb. B.D. mat'l}}$

Y - Expansion factor, dimensionless.

Y_1 - Water evaporated from feed based on temperature data, lb. / hr.

Y_2 - Water evaporated from feed based on dew point data, lb. / hr.

Y_3 - Water evaporated from feed below the feed port, lb./ hr.

$(C_p \mu / k)$-Prandtal number, dimensionless.

$(\mu / \rho D_v)$-Schmidt number, dimensionless.

β - D_2/D_1.

θ - Drying time per particle, hrs.

λ - Latent heat of vaporization at wet-bulb temperature, B.T.U. / lb.

μ - Viscosity, lb./ hr.-ft.

π - Average pressure in drying section, atm.

ρ - Density of gas stream, lb./ cu. ft.

ρ' - Density of outlet gas stream,,lb./ cu. ft.

ρ_s - Density of particle, lb./cu. ft.

ρ_s' - Density of final product, lb./cu. ft.

Subscript

f - Mean gas film condition.

FUNDAMENTAL STUDIES IN FLASH DRYING

by

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The purpose of this investgation was to obtain fundamental information for use in the design of floating bed type flash driers. The experimental data for the drying during the constant-rate period of particles of disc shape floated in a hot gas stream were obtained for a vertical drying section of a flash drier having a diameter of 16 inches for the first 7 feet and increasing from 16 to 18 inches in diameter over the remaining 8 foot of length. Unwaxed number 9 (48 mm.) milk bottle caps proved convenient for the drying study. Both temperature and humidity data were taken at various levels throughout the length of the drying section. These data were used to relate, so far as possible, the independent variables to the rates of mass and heat transfer during flash drying.

The total surface area available for mass and heat transfer is determined by the hold-up in the drier. This hold-up was dependent upon the feed rate, inlet and outlet densities or moisture contents of the particles, and the mass velocity of the gas at the top of the drying section. The hold-up in terms of the variables is given by the following equation:

$$\frac{F' \ (1 + W_1) \rho'}{H \ (1 + W_o)} = \frac{G'^{9.088}_2}{6.47 \times 10^{30}}$$

Where: F' = feed rate, lbs. /hr. - sq. ft. cross-sectional area at top of drier

ρ' = density of gas stream at top of drier, lb./cu.ft.

H = hold-up, lbs. product / cu. ft. drier volume

G'_2 = mass velocity of gas stream at top of drier, lb. / hr. - sq. ft.

W_o = moisture content of feed, lb. water / lb. bone dry material

W_1 = moisture content of product, lb. water / lb. bone dry material

The sensible heat necessary to carry on adiabatic evaporation of moisture comes only from the gas stream into which the moisture is being evaporated. The rate of evaporation was controlled by the mass and heat transfer characteristics of the gas film which surrounds the particle. These transfer characteristics depended primarily upon the "effective" gas velocity which was a function of the free-falling velocity of the particle. The effective mass velocity past the particles in the drier is given by the equation:

$$u' = 0.503\sqrt{d''\left(\frac{\rho_s}{\rho} - 1\right)}$$

where: u' = effective mass velocity past particle, ft./sec.

d'' = diameter of particle, inches.

ρ_s = density of particles, lb. /cu.ft.

ρ = density of gas stream, lb./cu.ft.

The values of the mass and heat transfer coefficients varied over wide ranges with changes in the mass velocity of the gas stream past the particles, the humidity of the drying gas, and the gas stream temperature. Experimental values of the j transfer factors which are expressions of the transfer rates in term of variables arranged in dimensionless groups. The j factors for mass and heat

transfer were related to a modified Reynold's number in experimentally derived equations. They are.

$$j'_d = \left(\frac{D_p G'}{\mu}\right)^{5.654} \Big/ \; 2.37 \times 10^{23}$$

$$j'_h = \left(\frac{D_p G'}{\mu}\right)^{5.654} \Big/ \; 1.82 \times 10^{23}$$

where: D_p = effective particle diameter $\sqrt{A_p/3.14}$, ft.

A_p = surface area per particle, sq. ft.

μ = average viscosity, lb./ ft.-hr.

G' = average modified velocity of gas stream, lb. / hr.-sq.ft.

$j'_d = (k'_G \; M_m \; P_f \; / \; G' \;) \; (\mu \; / \; \rho D_v)_f^{2/3}$

$j'_h = (\; h'_G \; / C_p \; G' \;) \quad (\; C_p \; \mu / \; k)_f^{2/3}$

k'_G = mass transfer coefficient, lb. moles/ hr.-sq. ft. - atm.

M_m = mean molecular weight of gas stream, lb./ mole.

P_f = film pressure factor, atm.

ρ = density of gas stream, lb. / cu. ft.

D_v = average diffusivity of air-water vapor mixtures, sq. ft. / hr.

h'_G = heat transfer coefficient, B.T.U./ hr. - sq. ft. - ^{o} F.

k = thermal conductivity of gas stream, B.T.U. - ft. / hr.-sq. ft. - ^{o} F.

Subscript f = mean film conditions.

The relationships presented are for a flash drier operated with a void fraction $\frac{\text{(volume of particles)}}{\text{(volume of drying section)}}$ of 99 percent or greater. No correction was made for the quantity of heat energy radiated from the hot walls of the

drying section to the particles being dried, and the transfer rates were obtained for the one particle size and shape only. Therefore, the use of these equations are restricted to estimation of the transfer rates for flash drying, during the constant-rate period, for disc or flaked material haveing a diameter of about 1-5/16 inches at a maximum gas temperature of 500° F. Other sizes and shapes of meterials being dried in this type of drier are expected to show a similar trend.