THE EFFECTS OF FEED PROPERTIES ON SPRAY DRYING

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A

THESIS

BY

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INTRODUCTION

Few of the numerous Chemical Engineering unit operations offer as many diversified and intriguing possibilities as spray drying. With a scope of applications already ranging over a wide section of the Chemical Industry, new, spectacular developments are constantly being discovered. Besides the obvious importance of its practical aspects, spray drying presents theoretical problems of challenging complexity. In spite of a large weight of experimental evidence, the theoretical development lags far behind the practical applications, owing principally to the large number of variables involved, and to their complex inter-relationships. To this should be added the fact that no adequate method has yet been found of assessing and controlling the value of such factors as turbulence, which undoubtedly play an important role in this unit operation.

F. H. Knelman (64) -- who was the first to investigate the relations of factors in spray drying in these laboratories -- pointed out that the study of spray drying was made particularly difficult, not so much because of a scarcity of fundamental data, but because of a marked lack of uniformity of presentation and definition, making a comparative study of the experimental evidence almost impossible. In an attempt to make such comparative studies possible, he proposed new relationships for efficiency and performance which, if generally adopted, would go a long way towards establishing a unifying approach to the problem. These relationships have been used in the present work to great advantage.

Since Knelman's historical review was first prepared, a fair number of significant papers on the topic have appeared in the literature, and have, of course, been included in the present literature survey. Aside from these additions and of a more critical approach to the subject in the light of recent developments, his general method of presentation has been retained in this review.

SPRAY DRYING AS A COMPLEX UNIT OPERATION

Spray drying may be defined as the drying of an atomized solution or slurry in contact with a stream of hot gases, under conditions permitting the recovery of the dried product.

Most drying operations for the production of a granular solid from a solution require at least three separate operations: bulk evaporation, drying of the solid, and grinding or pulverizing. In addition to which, units must be provided for the handling of the dry material, packaging, etc. Spray drying leads, in one operation, to a powdered product which may be packaged immediately, thus eliminating several of the steps required by the more conventional methods. A further advantage of this process is the extremely high drying rates, which permit the high temperature dehydration of heat-sensitive materials. These high drying rates are made possible by the intimate mixing of the sprayed droplets with the drying medium, and the high specific surface of the sprayed liquid. As long as the moisture content of the material is above the equilibrium value, its temperature will approximate the wet-bulb temperature of the drying medium, regardless of the dry-bulb temperature existing in the chamber. Once the product is thoroughly dried, its rapid removal into a cooler zone will protect it from excessive temperatures.

The operation of spray drying may be considered to consist of three main component parts: atomization of the liquid into fine droplets, drying of these droplets by intimate contact with the drying medium, and the separation and collection of the dried product. Each of these operations will be considered in detail.

I ATOMIZATION

One of the earliest problems to present itself to the designer of spray drying apparatus was the question of adequate atomization of the feed (99). Existing types of nozzles had been originally designed for other purposes and were not suitable for the process under consideration. The earliest type consisted essentially of a metal disk with a small orifice in the center, through which the liquid was forced at pressures up to 200 p.s.i. In this type of nozzle, the atomized liquid left the orifice in a solid cone pattern which prevented the intimate mixing of spray and drying medium which is so necessary for rapid drying. In addition, the small orifice became easily obstructed and tended to enlarge over a short period of time, thus requiring frequent replacements. Inadequacy of mixing of the two phases, necessity of frequent replacements, and variation in the degree of atomization due to enlargement of the orifice, all led to the design of special nozzles for spray drying, some involving entirely new principles of atomization.

(i) General Principles

The atomization of a liquid can be attained by one of two separate mechanisms. The first consists of forming the liquid into a sheet or film by means of the proper apparatus, and the subsequent deformation of the film into droplets. The deformation takes place as a result of the inherent instability of the liquid sheet, owing to the high velocity of the spray relative to that of the ambient medium in which it is atomized. The second method is characterized by the formation of unstable threads or filaments, instead of sheets or films, before it is deformed into droplets. The type of mechanism involved has a very large effect on the type of droplet formed. The film-forming atomizer tends to produce droplets in the shape of hollow spheres, due to the so-called "flag effect" (41). The unstable liquid film is whipped by the drying gases, much like a flag in the wind, so that it curls around itself and forms a liquid cylinder, which breaks up to form hollow droplets. The filament-forming mechanism, on the other hand, produces essentially solid droplets because in this case the thread is torn from the main body of the liquid without the formation of a hollow cylinder.

Owing to the high degree of dispersion of the sprayed droplets, the possibility of larger drop formation by collision is quite remote. The energy of formation of the spray must be provided by the spray mechanism, and Kleinschmidt (61) has divided the energy required into three main components:

(a) The energy required to form the surface of the spray against the surface tension. Mathematically, this is the product of the surface tension and the additional area created. Although the area of an atomized spray is considerable, the energy required for its formation is relatively small. Thus, Edeling has calculated that for an average drop size of 150 microns, although one litre of water will develop a cloud with a total area of 63 square meters, the energy of formation of the surface would be only slightly over one calorie (33).

(b) The energy required to produce deformation of the liquid. This is by far the greatest of the energy requirements, because of the enormous viscous forces to be overcome in the short time in which the liquid is exposed to the disrupting influence of the nozzle. This quantity is not easily computed and is best obtained from experimental data.

(c) The energy loss due to inefficient application of the available energy. This item obviously depends on the nozzle design and varies from a relatively low figure for an efficient film-forming atomizer to very high values for the filament-type spray nozzle.

(ii) Types of Nozzles

The most common classification of nozzles used in spray drying is based on the mechanical aspect of spray formation and usually distinguishes three main types: the pressure nozzle, the rotating disk type, and the gas-atomizing or two-fluid nozzle (56).

The most common type of pressure nozzle consists of a whirl chamber with an orifice at the outlet end. The energy required for atomization is fed to this nozzle directly as liquid pressure, which may be as high as 2500 p.s.i. The liquid feed under pressure enters the chamber tangentially and leaves the orifice as a widening conical film which rapidly breaks down into droplets in the spray. Variations of this are the impact type, in which the issuing jet strikes an obstruction beyond the orifice and at an angle to it, and the double jet type, in which two high pressure jets impinge on one another. Both of these produce a fan-shaped film of spray.

The rotating or centrifugal disk atomizer is, strictly speaking, not a nozzle. The liquid is fed to the center of a rotating disk or cup and is thrown off the periphery by centrifugal force. The dispersed liquid forms a horizontal sheet in the vicinity of the disk, and is subsequently broken up, as in the previous case, by the action of the turbulent gas stream. The energy required for atomization is commonly supplied by an electric motor which may be geared directly to the shaft of the disk. Speeds above 3600 and up to 20,000 R.P.M. are used. One advantage of this type is that no energy need be stored in the liquid for atomizing purposes. Thus it may be fed by gravity if convenient.

The third type of nozzle used in spray drying is the gas-atomizing or two-fluid nozzle. In this case, the liquid feed is broken up by impingement with a high velocity gas stream. The action of the compressed gas on the feed liquid is to tear off fine filaments or threads. These, in turn, disperse into droplets. In general, a higher pressure will produce a finer degree of atomization. The atomization energy is supplied as pressure in the disrupting gas stream, which may be as high as 100 p.s.i. for very fine atomization.

The above classification, based as it is on the mechanical device responsible for spray formation, is probably simpler, and certainly more common than that proposed by Fogler and Kleinschmidt (41), which depends on the physical principles involved, i.e. whether the spray is initially formed as a film or as a filament. Thus, according to their views, the various pressure nozzles and the centrifugal atomizer would come under the heading of film-forming types, while the gas-atomizing nozzle represents the second type.

(iii) Materials Handled

It is obvious that the amount of energy that can be stored in a unit mass of liquid by pressure alone is small compared to that possessed by a high velocity jet of compressed gases of equivalent amount. For this reason, the pressure nozzle is restricted to solutions of low viscosity which are easily broken up (41). The centrifugal and two-fluid nozzles, however, which do not depend on the amount of energy stored in the feed, can handle almost any solutions or slurries which can be pumped, although it may be convenient in certain cases to reduce the viscosity of the feed by raising its temperature.

(iv) Drop Size and Size Distribution

Since the size of the dried particle and the rate of drying are both dependent on the average size of the sprayed droplet, it is obvious that methods of determination and control of this variable are of vital importance to the process. This fact has been made particularly evident by the work of F. H. Knelman (64), who showed the close relationship existing between the drying rate and the droplet size for various operating conditions. Similarly, the distribution of the droplet diameters about the average value will affect not only the capacity of the spray drying chamber, but also the bulk density of the product and many of its physical properties. The average droplet size as well as the range of sizes produced will depend, in general, upon the method of atomization and on the energy input, as well as on the physical properties of the feed.

Most of the experimental data on droplet size and size distribution in a spray have been obtained with two-fluid nozzles. These nozzles give the finest atomization of any in use at the present time. As a matter of fact, the lower droplet size ranges are much too small to be used in spray drying, due to the difficulty of subsequently separating the dried particles from the drying medium. Fogler and Kleinschmidt (41) report that these nozzles tend to give a non-uniform droplet formation. They claim that high frequency vibrations, set up by the high velocity air stream, release dissolved gases in the form of fine bubbles in the droplets, and further create pressure pulsations which tend to alter the uniformity of the atomization. Furthermore, these authors claim that there is a large decrease in the temperature of the drops emerging from such nozzles, owing to the adiabatic expansion of the gas stream, which will affect the atomization to some extent. Using the well-known expression for the temperature drop during isentropic expansion of an ideal gas, they have tabulated values showing considerable cooling effects in the nozzle, for both air and steam as atomizing fluids and under various conditions. However, these calculations assume reversible conditions, which certainly do not prevail in this case, and have never been supported by experimental evidence. Moreover, their views regarding the lack of uniformity of drop size distribution are not supported by the work of Houghton (56), who has presented evidence that the size distribution of a two-fluid nozzle is much narrower than that produced by a pressure nozzle of the same capacity.

Nukiyama and Tanasawa (91) have developed an empirical equation expressing the droplet diameter in terms of the physical properties of the feed and the rates of flow of the two fluids:

$$D_o = \frac{585}{v}\sqrt{\frac{\sigma}{\rho}} + 597\left(\frac{\mu}{\sqrt{\sigma\rho}}\right)^{0.45}\left(\frac{1000\,Q_l}{Q_a}\right)^{1.5} \qquad (1)$$

where: D_o = the diameter of a single drop with the same ratio of surface to volume as the total sum of drops in microns.

v = relative velocity between the air stream and the liquid stream, in meters per second.

Q_l/Q_a = ratio of the volumetric rate of flow of the liquid to that of the air at the "vena contracta", or throat, of the nozzle.

ρ = liquid density, in gm/cc.

μ = liquid viscosity, in poises.

σ = liquid surface tension, in dynes/cm.

This equation was developed specifically for two-fluid nozzles. It will be noted that it is not dimensionally consistent, and the constants, therefore, will vary with the system of units used.

Data on distribution of drop sizes have been correlated by these authors in another empirical equation:

$$\frac{dn}{dx} = a\,x^{p}\,e^{-bx^{q}} \qquad (2)$$

where: x = the diameter of an individual drop, in microns.

n = the number of drops with diameters between zero and x in the total sample.

a, b, p, q = constants.

The value of p was found to be 2 for all cases studied by Nukiyama and Tanasawa. Lewis et al (77) found that q = 1 if the velocity and volume rate of the air stream are high. Using these values, and converting equation (2) into log form, a more convenient equation is obtained for the analysis of experimental data:

$$\log_{10}\left(\frac{1}{x^2}\frac{dn}{dx}\right) = \log_{10} a - \frac{bx}{2.3} \qquad (3)$$

A plot of the left hand term against x should yield a straight line of slopeb/2.3. D_o can then be calculated by the use of published tables (77). The value of q gives an indication of the distribution of the particle sizes, a high value representing a narrow range and vice versa. These equations have been tested by Lewis et al (77), and have been found to apply to their own data as well as those of other workers (56, 73, 106). Equation (1) has been found to fit the experimental results particularly well in the range: ρ = 0.7 - 1.2 gm/cc, σ = 19 - 73 dynes/cm, μ = 0.003 - 0.5 poises, and v less than the speed of sound (77).

Equation (1) indicates that as the relative velocity increases at the throat of the nozzle, D_o approaches a definite minimum value. This has been confirmed by Knelman (64) using this equation, and by Edeling (33) from theoretical considerations. Experimental work by Sauter (104) leads to the same conclusion. Micrographic studies have further shown that as the relative velocity increases, more and more droplets of the minimum size are formed, thus narrowing the droplet size distribution (89, 90). These considerations indicate that there is a certain pressure above which any increase will not result in a corresponding decrease of average droplet size.

In considering the practical applications of equation (1), it should be noted that D_o is an average diameter, defined as the diameter of a drop with the same surface to volume ratio as that of the total spray cloud. It does not represent the minimum size formed, nor is it associated with the degree of uniformity of the drop sizes. Furthermore, the accuracy of the expressions presented by Nukiyama and Tanasawa is limited by a number of complicating factors, the effects of which are difficult to assess. Heat transfer between the two fluids in the nozzle, possible evaporation of the liquid in the nozzle chambers, and the difficulty of estimating the physical properties of the two fluids under the conditions prevailing at the throat all contribute to the complexity of the problem.

Considerably less work has been done on pressure nozzles and centrifugal disks than on two-fluid nozzles. Houghton (56) presents the general manner in which the drop size varies with pressure and the physical properties of the feed in a pressure nozzle. Lewis et al (77) have modified the Nukiyama and Tanasawa equations to apply to pressure atomization as well, but these modifications cannot be extended to the centrifugal disk. In the latter type, the particle size tends to decrease with increasing rotational speeds, decreasing viscosity, and decreasing discharge rates (56). Walton and Prewett, in a recent investigation (119), found that the degree of uniformity of atomization decreased as the feed rate increased, at a constant speed of rotation of the disk. They postulate that, at low feed rates, the drops are formed and released individually from the edge of the disk. The centrifugal force and the surface tension are almost equal until the drop breaks away. These views are supported by the narrow size distribution range obtained at these flow rates. Further proof is contained in the data of Phillip (94), who found that solid dried particles are produced from these sprays at low feed rates. At high feed rates, the centrifugal forces are aided by flow inertia forces, and the liquid leaves the disk as a sheet, as suggested by Fogler and Kleinschmidt (41). Walton and Prewett present the following equation for estimating the drop diameter. It has been found to apply well if low feed rates are used, but is not applicable if the liquid leaves the disk in sheet form:

$$D = \frac{3.8}{\omega} \left(\frac{\sigma}{D_d \rho} \right)^{\frac{1}{2}} \tag{4}$$

where: D = diameter of the drop, in cm.

D_d = diameter of the disk, in cm.

ω = angular velocity of the disk, in radians/sec.

σ = surface tension of the liquid, in dynes/cm.

ρ = density of the liquid, in gm/cc.

For two-fluid nozzles, a drop size of 15 to 70 microns is the average for commercial installations. 15 microns is about the lower average limit for water according to equation (1). Sauter (105) found 6 microns to be the minimum drop diameter obtainable from a two-fluid nozzle. The operating range of the other nozzles discussed is about 50 to 300 microns for the average droplet diameter.

(v) Relative Performance

Of the three types of nozzles described, the most expensive in both capital and operating costs is the centrifugal disk type. However, if good control of atomization is required over a wide range of feed viscosities, as well as large capacity, the cost may be justified. Maintenance on such a type is light and replacements infrequent. Two-fluid nozzles are cheaper than centrifugal atomizers and will handle about the same range of viscosities. Their capacity is limited, however, and they are not in general use outside of experimental equipment. Pressure nozzles require less power than either of the other types but operating expenses are increased by frequent shut-downs for repairs or replacement of the nozzle. Their individual capacity is less than that of the centrifugal type, but the volume of liquid sprayed can be increased by connecting two or more nozzles in parallel in the drying chamber. In general, the pressure nozzle is not as flexible in operation as the other types, nor does it handle as wide a range of viscosities. Its low initial cost, moreover, is somewhat offset by the cost of the high pressure pump and piping necessary for its operation.

II DRYING AND THE SPRAY CHAMBER PROCESSES

The main difference between spray drying and other forms of drying resides in the atomized condition of the feed. This difference dominates every aspect of the process: it affects the heat and mass transfer rates, the design of the equipment, the method of powder collection, the physical properties of the dried product, etc. In addition, the theoretical considerations commonly met in the older methods of drying are complicated by the introduction of the problems of flow conditions and turbulence, and their effect on inter-phase transfer rates.

The drying of an atomized droplet, just as in the case of a bulk wet solid, is essentially a problem of mass transfer across a fluid film. In both cases, the driving force is the difference between the vapor pressure at the surface and that of the drying medium, but the very nature of spray drying operations introduces new complexities which are not found in the more conventional methods. For pure water or colloidal suspensions, the driving force will be about constant during the drying period. In the case of solutions, however, the vapor pressure at the surface of the droplet will decrease as the solution becomes more and more concentrated. This decrease of vapor pressure with increase in concentration will not be considered in the following discussion because it greatly complicates the treatment, and because many spray drying operations deal with dilute slurries or colloidal suspensions in which this phenomenon does not occur.

Before a study of the drying processes proper can be made, considerations must first be given to the path assumed by the droplet in the chamber.

(i) Droplet Path

The path taken by the droplets as they emerge from the nozzle is a very important factor in the design of spray drying equipment.

This path may be considered as consisting of two distinct parts: the initial trajectory traversed by the droplet as a result of the impetus received at the nozzle; and the final path taken under the influence of the turbulent drying gas. Thus it can be seen that the droplet path can be controlled, in a large measure, by the correct design of the spray nozzle during the first stage of its course, and by that of the chamber during the second.

The droplets emerging from the nozzle possess high velocities, of the order of several hundred feet per second. The distance travelled is proportional to the momentum received at the nozzle. Thus, a droplet 100 microns in diameter will travel about two feet from the nozzle, while a 300 micron droplet may go as far as six feet before it reaches its terminal velocity (33). Smaller drops will of course be restricted to shorter distances for the same conditions. The time required for a droplet to reach its terminal velocity is measured in fractions of a second, and seldom exceeds 0.1 second for the largest drops (33). The average distance of travel for a droplet under the conditions commonly met in practice has been estimated to be about one foot from the nozzle in a time of approximately 0.01 second (111).

While the droplet is travelling under the influence of the momentum received at the nozzle, it is not affected appreciably by the surrounding medium. However, as it approaches its terminal velocity, the effect of the turbulent gas stream increases until, when the relative velocity of the spray and gas streams drops to zero, the droplet assumes the path of the drying medium. In most commercial dryers, the hot gas is admitted tangentially and follows a circular path, having minimum velocity at the vertical axis of the chamber and increasing toward the wall. The maximum tangential velocities vary from about 4 to 20 feet per second (111).

Lapple and Shepherd (72) have derived equations of particle motion for several cases, suitable for the calculations of particle trajectories. According to Marshall and Friedman, however, these are not strictly applicable to spray drying because they were derived by consideration of a single droplet (83). In addition, the equations presume a constant drop diameter, which is not realized in spray drying. Turbulence in the gas stream and eddy currents would further complicate the calculations. However qualitative these considerations may be, they throw considerable light on the complex processes taking place in a spray drying chamber.

In its simplest aspects, the system of forces acting on a droplet conveyed by the hot gas medium may be reduced to three: the gravitational force, which tends to make the particle descend; the frictional force, which gives it its tangential velocity; and the centrifugal force, which causes it to move outward toward the wall. The following equations express the magnitude of these three forces mathematically:

Gravitational force: $$F_g = \pi g D^3 (\rho_s - \rho)/6 \tag{5}$$

Frictional force: $$F_f = \pi C u^2 D^2 \rho/8 \tag{6}$$

Centrifugal force: $$F_c = \pi D^3 (\rho_s - \rho) u^2/6r \tag{7}$$

where: D = droplet diameter, in cm.

ρ_s = actual droplet density, in gm/cc.

ρ = gas density, in gm/cc.

C = dimensionless drag coefficient, a function of the Reynolds number.

u = relative tangential velocity between the droplet and the gas stream, in cm/sec.

r = radius of curvature, in cm.

The capacity of the drying chamber is determined mainly by the requirement that no sprayed particle shall hit the walls when wet, under operating conditions. It will be noted from equations (6) and (7) that a large droplet will be affected by the turbulent gas stream to a greater extent than a small one, due to the large D terms. Similarly, a large droplet will take on a greater momentum at the nozzle than a small droplet with the same velocity. It follows that proper chamber design should be based on the largest droplet diameter in the given spray: if these are properly dried, without accumulation of wet particles on the walls, then the smaller particles will have obviously received the same treatment. As a corollary, the capacity of a given chamber can be increased by decreasing the average drop size, or by reducing the size distribution about the average value.

(ii) Drying

Basically, there is little difference between drying and evaporation. In the chemical industries, however, evaporation is usually associated with the removal of large quantities of water by vaporization, while drying concerns the removal of a lesser amount of water from a wet solid. The two stages of drying -- the constant rate and the falling rate periods -- occur in the dehydration of a dispersed spray just as in the case of a bulk solid. The rate of removal of moisture during the constant rate period is governed by the rate of mass transfer through the laminar film of gas surrounding the droplet. Since the rate at this time is comparable to that from a pure water surface under the same conditions, there is no line of demarcation between the evaporation period and the first step of the drying process. The difference is one of definition only.

From the point of view of spray drying, the most interesting and most important step in the process is the falling rate period. This may be further subdivided into two steps: the period of unsaturated surface drying, where the diffusion of vapor through the gas film still controls; and the final stage, where drying is now controlled by the rate of internal diffusion of the liquid or vapor. D. A. Smith (111) believes that the period of unsaturated surface drying is negligible in spray drying because of the uniformity of the surface of the spherical particles formed. The existence of the internal diffusion drying period, on the other hand, is proven clearly by the formation of hollow spherical particles, often punctured or fragmented, as the dried product. In the falling rate period of drying, evaporation proceeds from a continuously receding spherical liquid surface. During this last stage and as the surface of the particle dries, the rate of heat input will increase, and will become greater than the rate at which the heat is removed by diffusion of the vapor. The formation and expansion of the vapor inside the particle will cause it to expand if it is a plastic substance, and may eventually cause it to rupture. Alternatively, if the nature of the substance is such that the necessary plasticity is absent, the dried particle will exhibit marked shrinkage effects or surface checks during the same period. It is probable that fragmentation of the particle during the last stage of drying, which proceeds at a very slow rate, will appreciably shorten the time necessary for the particle to be in contact with the hot drying medium. Ladish (66) has experimented with addition agents to increase drying rates and thus allow the use of lower inlet temperatures. He found that a few percent of a volatile solvent would cause the particle to explode, and thus permit complete drying at a lower temperature level than normally required. A similar study has recently been reported by Chu, Stout and Busche (25). These authors found that addition of varying amounts of methanol or acetone to a solution of Santomerse No. 1 first caused the bulk density of the spray dried product to decrease to a minimum value at a solvent concentration of 15 to 20 percent, and then to increase with higher solvent concentrations.

Very little data is available on the actual time taken for a particle to dry. This is unfortunate because such data would enormously simplify the design of the spray chamber. The reason for the lack of information is probably to be found in the experimental difficulties encountered in any approach to the problem. Various theoretical equations have been advanced (83), but they are restricted by too many limiting assumptions to be of practical value. The overall drying time, including separation and collection of the product, has been found to vary from 20 seconds (41) to 45 seconds (49). In any event, this information is of little value. Edeling (33) has reported drying times ranging from 3 seconds for 100 micron to 6.3 seconds for 300 micron particles. The material used originally contained 80 percent of water and did not shrink on drying. When shrinking occurred, a 20 percent increase in the drying time was reported, owing to slower drying rates during the falling rate period.

With adequate insulation, the operation of a spray dryer will tend to approach adiabatic conditions. The decrease in temperature and the increase in humidity will follow approximately the adiabatic cooling lines on a humidity chart. Maximum utilization of heat will take place when the exit gases leave the chamber saturated (at the wet-bulb temperature).

(iii) Mass and Heat Transfer

During the drying of a wet particle, two processes are taking place: the transfer of mass away from the particle and the transfer of heat into it. If the heat transfer is considered as a diffusion of 'hot' molecules to a cooler area, it will be seen that there is a remarkable similarity between the postulated mechanisms of heat and mass transfer. There has been a growing tendency in recent years to regard the three processes of heat, mass and momentum transfer as different aspects of the same operation -- diffusion, and to develop a basic theory to include all three.

Chilton and Colburn (24) first suggested that data on mass transfer be correlated in terms of the dimensionless group j_D:

$$j_D = \frac{K_c}{U_o} \frac{P_{BM}}{P} \left(\frac{\mu}{\rho D_v} \right)^{2/3} \tag{8}$$

A similar group, j_H, was defined for heat transfer:

$$j_H = \frac{h}{CU_o \rho} \left(\frac{c\mu}{k} \right)^{2/3} \tag{9}$$

where: K_c = mass transfer coefficient, in gm. moles/(sec.)(sq.cm.) (gm. mole/cc.)

U_o = average velocity of the gas stream, in cm./sec.

P_{BM} = log mean pressure of inert gas, in atm.

P = total pressure, in atm.

μ = viscosity of the gas, in poises.

ρ = density of the gas, in gm./cc.

D_v = diffusivity, in sq. cm./sec.

h = heat transfer coefficient, in cal./(sec.)(sq.cm.)(°C).

c = specific heat of the fluid, in cal./(gm.)(°C).

k = thermal conductivity, in cal./(sec.)(sq.cm.)(°C/cm.).

By means of these correlations, the authors found it possible to predict mass transfer data from available data on heat transfer. T. K. Sherwood, in an article on mass transfer and friction in turbulent flow (110), traces the development of the analogy between the three different diffusional processes from Stefan's Law to the complex expressions now being advanced.

Hougen and co-workers (47) investigated mass, heat and momentum transfer through granulated solids, using spheres and cylinders with a range of diameters of 0.2 to 0.4 inches. Their data, plotted as j_D or j_H against Re, fell on a smooth curve throughout the entire range of values of Re. They reported that the ratio of j_H to j_D was constant at 1.076 for all conditions of flow. T. K. Sherwood, in a review of their work, agreed that j_H/j_D was probably close to unity, but argued that the experimental value obtained was inaccurate due to certain assumptions made and the use of a humidity chart for the calculations. He also suggested that the correlation of j_D and Re was too high in the viscous flow region. Subsequent investigation by Wilke and Hougen (120) proved this to be the case. The various correlations reported for j_D and j_H vs. Re are in fairly good agreement when the different experimental conditions, shapes of particles used, etc., are taken into account (47, 54, 79, 120, 121).

Chilton and Colburn, in their derivation of j_D and j_H, found these functions to be equal to $\frac{1}{2}f$, where f is the skin friction. This has been verified by Maisel and Sherwood (79) for cases where the skin friction can be determined. There is at present no correlation between j_D and f for spheres because the proportion of the total drag due to the skin friction cannot be calculated. Lapple and Shepherd (72) report that for spherical particles, the total drag coefficient, $C = 24/Re$ in the viscous region and that C is constant at 0.44 in the turbulent region.

It is expected that continued investigations will reveal similar correlations as being applicable to spray drying. For these, the continually changing diameter of the droplet will have to be taken into account, as well as the changing velocities and the falling rate period of drying.

Edeling (33) has presented an equation relating the coefficient of heat transfer to the amount of water evaporated during the free fall of small spherical droplets:

$$h = 0.22 W \qquad (10)$$

The author justifies the high values of h obtained by this equation on the grounds that drying proceeds faster from spherical surfaces than from flat surfaces under similar conditions. It is probable, however, that the high drying rates noted in spray drying are due to the large surface exposed rather than to abnormally high values of h, since the effect of curvature on vapor pressure is negligible above a droplet size of 0.01 micron.

(iv) Mixing and Turbulence

It has been calculated that evaporation of one percent of the water content of a droplet will form a cloud of vapor approximately seventeen times its volume (41). If the droplet is suspended in still air, this vapor will form a cool area about it which has a radius about five times that of the droplet. This fact dramatically emphasizes the need for turbulence in the drying zone, to provide continuous removal of the cool vapor. Once the vapor has diffused through the laminar film surrounding the droplet, its further transfer across the turbulent gas stream proceeds mainly by the mechanism of eddy diffusion. Maisel and Sherwood (79) have recently suggested that the degree and scale of turbulence in the gas stream might influence the rate of drying by decreasing the thickness of the laminar film.

The turbulence should be carefully controlled, since stray eddy currents can cause wet material to hit the walls, or to collide and form aggregates which will not dry completely in the time available. It would appear that counter-current flow would provide maximum mixing of spray and air. However, this causes the dried particles to come in contact with the hottest portion of the air, which may damage heat-sensitive materials. Counter-current flow is also restricted by the conveying velocity of the particles (41).

(v) Chamber Design

The design of the spray chamber is determined by the path to be followed by the drying particles. The general criterion is that no wet or partially dried material shall hit the walls of the chamber. Thus the diameter required will depend to a great extent on the initial velocity of the spray relative to the drying medium. The height will be a function of the path of the drying gases. Many other factors must be taken into account, however, such as the type of nozzle used, the manner in which the hot gases are to be introduced into the chamber, the temperature range to be used, etc. The chief variation in the designs of drying chambers in use today depends on the initial relative directions of spray trajectory and drying gases.

The early types of spray dryers used horizontal, rectangular drying chambers, the feed being sprayed in at one end. The Merrel-Soule (180, 181) and the Fleischer (142) dryers are examples of this type. In recent years, however, the trend has been toward the design of vertical chambers, the spray and gas streams being concurrent and downwards. Depending on the relative directions of flow of spray and gas, D. A. Smith (111) has classified the common designs into five general types:

(a) The mixed flow type, in which the gases initially travel concurrently with the spray, and then reverse their direction and flow up the center of the chamber counter-currently. This method of handling the gas increases the thermal efficiency of the dryer without subjecting the partially dry material to damaging temperatures. The Gray-Jensen dryer employs this technique (138, 148, 149).

(b) The horizontal concurrent type of drying chamber, mentioned above.

(c) The vertical, up-flow counter-current type. In this design the spray enters at the top and the gases at the bottom of the chamber, the product being collected at the bottom. The system is restricted to low operating temperatures, or to materials not affected by heat. The Ravo-Rapid spray dryer, used in Europe, employs this method for the drying of soaps, etc. (94).

(d) The vertical, concurrent down-flow type is a very widely used design. The gas enters tangentially to the spray at the top of the chamber, and the dried particles are carried out in the exit gas at the bottom. The lower part of the chamber may be conical to increase the velocity of the hot gases. The Dickerson tower type (162, 168) and the Mojonnier dryer (49) are representative of this category.

(e) The vertical, up-flow concurrent type, in which the spray and gas are admitted at the bottom of the chamber. This chamber acts as a classifier, the larger particles being removed at the bottom, while the fines are carried out at the top by the spent gases.

These five types of drying chamber form the basic designs in commercial use. Many modifications developed, however, as the need arose. In some dryers (41) cold air can be admitted about halfway down the chamber to chill materials which are sticky above room temperature. The Bowen spray dryer (15) divides the entering gas into two streams. One stream enters the chamber from an annular space surrounding the nozzle and the other enters tangentially through long narrow openings. The result is to enclose the spray in a dynamic cylinder of hot gases which effectively prevents any wet material from reaching the walls or roof. This design reduces considerably the size of the chamber needed for a given capacity. The Peebles dryer(189) also uses a method of controlled turbulence. In this case, both streams of air enter tangentially, but in opposite directions. The result is that the spray is surrounded by two concentric air streams, revolving in opposite directions.

Most dryers have stream and water connections to the feed lines for cleaning and for fire control. It is also common practice to begin and end the drying operation with water to avoid the production of material which does not meet the specifications required. The chamber should be so designed that repairs and maintenance can be quickly and efficiently carried out. The nozzle, in particular, should be easily accessible, and inspection ports should be provided at points such that any part of the chamber can be seen. Some dryers are operated under a slight vacuum to allow the opening of these ports without loss of material while the dryer is in operation.

(vi) Effects of Operating Variables

The effects of certain operating variables on the capacity and efficiency of the drying chamber were investigated extensively for the first time by F. H. Knelman (64). By varying one operating condition at a time, and holding all others constant, he studied the effects of the drying gas rate and temperature, the depth of the nozzle below the chamber roof, and the temperature of the inlet feeds, which consisted of pure water and of a colloidal solution of fixed concentration.

He found that, as the variable was increased, the capacity in every case rose to a maximum and then decreased. By choosing a proper combination of operating conditions, an optimum chamber capacity could be realized. The drying thermal efficiency varied in the same manner as the capacity, the two maxima occurring at the same point if the operating temperatures were constant. When varying the temperature of the drying medium, it was found that the efficiency reached a maximum at a lower temperature than for maximum capacity.

An interesting effect reported by Knelman was the fact that the optimum nozzle position was quite critical and was independent of the rate of flow of the drying gases, within the range studied. He noted, however, that the optimum nozzle position was much lower in the chamber for the colloidal solution than for pure water. Another unexpected result was the fact that, as the inlet feed temperature was increased, the capacity increased also, reaching a maximum when the feed entered at the wet-bulb temperature of the drying medium. The reason for decrease in capacity at still higher temperatures has not been adequately explained, and there is some doubt as to the validity of these data. Owing to lack of information concerning the effects of feed properties and concentration, Knelman found it impossible to formulate correlations between the various variables nor to establish a preliminary theory of design, based on the experimental results obtained.

III COLLECTION OF THE PRODUCT

The third and last of the principal spray drying operations is the separation of the dried material from the spent gases. Although some dryers may act as classifiers (49, 94, 148), there is always some air-float material which must be removed before the gases are exhausted to the atmosphere or partially recycled. The methods of product collection vary, depending on the type of dryer, the nature and value of the product, and other factors. The usual procedure is to remove the main portion of the product in a cyclone separator and to follow this step with a second separation stage if necessary. This stage normally consists of a device such as a fabric filter, bag collector, scrubbing tower or preconcentration chamber.

If the bulk of the product is collected at the bottom of the chamber, as in the countercurrent design where the chamber itself acts as a separator, a conveyor is usually installed to carry it to the packaging stage. In this case, the spent gases carry away the fines, which are removed by one of the devices mentioned above. In many dryers the product is conveyed directly to the collector by the exhaust gases. In this type the drying chamber is usually cone-shaped, tapering to the diameter of the exhaust duct, thus giving the gas stream the necessary high conveying velocity. Use of a secondary collecting device, following the main cyclone collector, will be dictated by the cheapness of the product, by its nuisance value in the atmosphere, and by the economics of the specific process. The higher the collection efficiency, the higher the pressure drop across the equipment, with a corresponding increase in the power costs.

The efficiency of single cyclone collectors is rather limited, and they are currently being replaced by the more efficient multi-tube design. Collection efficiencies of 98 percent and better have been reported for particle sizes above 9 microns. Fabric filters are not extensively used due to the damaging effects of the high temperature gases on the fabric, and to the fact that the product may be wetted by condensation from these gases (41). One of the most efficient ways to reclaim the air float materials is to pass the exhaust gases through a preconcentration chamber (148, 149). The feed is sprayed into the exhaust gases, reclaiming both the fines and some of the waste heat, and then proceeds to the drying chamber in a more concentrated condition.

The type of collection equipment to be used must be based on considerations of the drying process. The type of drying chamber, the nature and value of the product, the temperature range used and the local cost of power all play a part. Only an economic balance will indicate how extensive the collection equipment should be.

PRACTICAL ASPECTS OF SPRAY DRYING

I HEATING OF THE DRYING GAS

The heating medium must be raised to a temperature which may vary from 300 to 1400 degrees F., the latter being the upper limit for most materials of construction used in spray drying. This may be done in a direct heater, where the products of combustion pass into the gas stream, or indirectly, by means of a heat exchanger.

When a direct heater is used, the drying medium consists of a mixture of air and combustion gases. Care must be taken that the gases are clean and will not contaminate the product. This may be accomplished by burning only clean grades of coal or oil, or by using gaseous fuels such as natural gas. Soot or ash may be filtered or scrubbed out if necessary. Direct heating may be unsuitable for the drying of some food products if the flavour is easily impaired (49).

Indirect heaters are used where it is necessary to avoid contamination of the drying medium. In choosing such a heater, consideration should be given to the temperature level of the heat required. Low pressure steam, although often available at low cost, is limited to very low temperature levels. High pressure steam or superheated steam may be used for higher temperatures, but their use is expensive (41). The heat may be transferred directly to the exchanger by means of combustion gases from a furnace or by means of oil or diphenyl for the higher temperatures. Electrical heating is limited in most localities by the cost of power, and has been restricted to experimental dryers (122).

II NATURE AND PROPERTIES OF SPRAY DRIED MATERIALS

One of the main advantages of spray drying resides in the fine granular form of the product. Knelman has made an extensive study of the various factors affecting the particle size and came to the important conclusion that atomization exerts the largest single influence: under identical spray and drying conditions, the large initial droplet gives the larger dried particle (64). As a corollary, he showed that, everything else being constant, perturbances in the formation of the particle during the falling rate period of drying, such as shrinking, surface checking, case-hardening, expansion into a sphere, bursting, etc., will occur to the same extent irrespective of the initial droplet diameter. However, prediction of the final value of the average particle diameter, D_p, can only be made if a knowledge of the specific process occurring during the falling rate period has been previously obtained. Such knowledge is, unfortunately, largely empirical at the present stage of development. Of lesser importance is the influence of the operating variables: an increase in the inlet feed temperature caused progressive expansion of the dried particle, while a progressive shrinking was observed at higher volumetric rates of drying gases. Finally, microscopic studies showed that the characteristic particle from a two-fluid nozzle was spherical in shape, hollow and transparent. Higher drying rates often resulted in rupture of the spherules into egg-shell shaped fragments.

According to the analysis made by Fogler and Kleinschmidt (41), particles obtained from pressure nozzles exhibit the same appearance and properties as these obtained by Knelman from two-fluid nozzles.

T. B. Phillip (94) studied the nature of the particles formed from a centrifugal atomizer. Using a solution of low surface tension, he found that at low atomizer speeds, the particles formed were solid and spherical, the diameter decreasing as the speed increased. Above a critical speed, however, the particles became hollow, and the diameter increased with speed until a bursting point was reached. Solutions of different surface tensions or viscosities may show only a portion of this range of properties. This evidence supports the mechanism of spray formation from centrifugal disks put forward by Walton and Prewett (119).

The control of the bulk density has a large bearing on the economy of the spray drying operation, particularly in its effect on product collection and on packaging costs. A drop in the density by a factor of one-half would double the size of the package necessary to contain a given weight of the material. For this reason, the production of hollow spheres as dried particles is in some cases disadvantageous, as it often leads to a low bulk density. This tendency may prove prohibitive in certain cases -- such as in the drying of gelatin -- since the final material encloses too much air and results in frothy solutions when redissolved.

D. A. Smith (111) summarized qualitatively the relationship between bulk density and the various operating variables: the density varies directly with the feed concentration, the inlet feed temperature, and the drying rate. It is inversely proportional to the inlet drying gas temperature, the average drop size, and the uniformity of atomization. Knelman verified the correctness of these views experimentally, at least in the variation of bulk density with inlet gas temperature, average drop size, inlet feed temperature and rate of drying gas (64). Further support for the findings of Smith and of Knelman has been supplied by Chu, Stout and Busche, who investigated the variation of bulk density with inlet gas temperature, degree of atomization and concentration of the feed solution (25).

Little data is available concerning the equilibrium moisture content of spray dried powders. Greene et al (49), in a discussion of the production of spray dried egg powder, found that the stability of the product increased as the final moisture content was reduced, as little as 5 percent being detrimental to the powder. Consideration is given in this paper to various factors affecting the reduction of the moisture, and to special dryers designed to yield a bone-dry product. To reach very low moisture levels, two stages of drying are recommended, the first stage being spray drying, and the second a flash drying operation, using dehumidified air in a closed system.

From their studies of the spray drying characteristics of Santomerse No. 1, Chu, Stout and Busche report that the moisture content of the product decreases slowly as the drying temperature is increased, but is virtually independent of the pressure of the atomizing air fed to a two-fluid nozzle, and of the concentration of the feed solution (25). These authors further found that the addition of small amounts of organic solvents to the feed solution at first caused a marked decrease in the moisture content of the product, but as the amount of solvent was increased, the moisture content increased until, at a solvent concentration of 50 percent, the product contained about the same amount of moisture as that dried with no volatile materials added.

Due to the large surface area per unit weight of spray dried materials, their solubility is increased above that of products dried by other means. In addition to solubility, other properties are altered somewhat. Spray dried foodstuffs are usually more stable and will keep longer than those dried by other methods. There are no general correlations of the various data available. The properties of digitalis have been improved by spray drying (9); the use of addition agents to improve the stability of various products has been tried (49, 66); and techniques for improving the quality of dried eggs have been reported (49).

The control of the properties of the product rests mainly on qualitative empirical data. Much work has yet to be done on this topic.

III EFFICIENCY OF SPRAY DRIERS

The thermal efficiency of a spray dryer is largely dependent on the extent to which the drying gases approach a condition of equilibrium with respect to the water vapor pressure of the drying material, the theoretical maximum being obtained when the exit gases leave saturated. This condition of saturation is seldom achieved in practice, and in many cases would prove undesirable due to the possibility of condensation of the water vapor on the dried product in the relatively cool collection system. The efficiency is lowered somewhat by heat losses to the surroundings, but this effect becomes small as the chamber diameter is increased.

Water will evaporate from the droplets and partially dried particles as long as the partial pressure of the vapor in the gas stream is less than the vapor pressure of the liquid. Thus a higher theoretical efficiency would be expected for a suspension than for a solution, because the vapor pressure of the solution continually decreases as it becomes more concentrated which, in turn, would raise the equilibrium temperature of the exit gases. In commercial dryers, the exit temperatures are considerably above the equilibrium temperature, because, below a certain temperature, the transfer of heat to the particles becomes too small to produce the required evaporation in the time available (99).

The efficiency of a spray dryer is dependent on many factors. It is directly proportional to the inlet gas temperature, but in many cases this temperature is restricted by the nature of the product. Decrease in efficiency due to radiation loss to the surroundings was mentioned above. Finally, the efficiency will depend on the degree of utilization of the heat contained in the inlet gases, which, in turn, depends on the chamber design. The efficiency may be increased to some extent by recovery of some of the heat in the exit gas stream, the most successful methods being by partially recycling the gas (94, 189), and by using the exhaust gases to preconcentrate the feed (94).

In order to compare the performance of different installations, a standardized method of determining the theoretical efficiency should be used. Fogler and Kleinschmidt suggest the use of the expression:

$$E = \frac{\left(1 - \frac{R}{100}\right) \left(T_1 - T_2\right)}{T_1 - T_o} \qquad (11)$$

Knelman, however, pointed out that this expression gives no indication of the heat actually used for the evaporation of the water, nor will it give an efficiency of 100 percent, even when the exit gases emerge saturated from the chamber. He accordingly developed the following equation, which he termed the 'drying thermal efficiency' of the chamber:

$$E = \frac{W \lambda_{WB} \times 100}{G(T_1 - T_{WB})\, S_1 + W(T_F - T_{WB})\, S_F} \qquad (12)$$

This expression is based on the wet-bulb temperature of the drying gases as the datum and will therefore indicate an efficiency of 100 percent at the optimum point. Knelman also defines the 'heat economy of the drying chamber' by means of the same expression, replacing T_{WB} by the temperature of the surroundings, T_d.

Symbols: E = Thermal efficiency of the spray chamber, percent.
R = Radiation loss, percent of total temperature drop in drying zone.
T_1 = Inlet drying gas temperature, in °F.

T_2 = Exit gas temperature, in °F.

T_o = Datum, temperature of surroundings, in °F.

T_f = Inlet feed temperature, in °F.

T_{WB} = Wet-bulb temperature, in °F.

W = Rate of evaporation of water, in lb/hr.

λ_{WB} = Latent heat of vaporization of water at the wet-bulb temperature, in B.t.u./lb.

G = Weight rate of flow of drying gases, in lb/hr.

S_1 = Humid heat of entering gases, in B.t.u./(lb)(°F).

S_F = Specific heat of entering feed, in B.t.u./(lb)(°F).

IV APPLICATIONS OF SPRAY DRYING

Spray drying is widely used in the process industries, and its scope of application to specialized operations is constantly increasing. Its peculiar use for the dehydration of heat-sensitive materials has already been emphasized. It is employed in the drying of insecticides and allied materials (1), and is particularly suitable for poisonous products because no manual handling of the dry product is necessary until it is safely packaged. Its use in the food industry is widespread: Milk (3, 35, 154), eggs (49), soluble coffee (111), fruit and vegetable juices (98), potato powders (19), and many others are being dried successfully by this process.

The pharmaceutical industry uses spray drying to a large extent. Taylor (114) reports a method by which acetone is evaporated from a water-penicillin mixture. Penicillin and blood plasma have been successfully spray dried at a lower cost than by the conventional sublimation process (4). Powdered soap and detergents are produced by the same method (86, 97). Some materials such as molasses and glucose which are sticky when wet, cannot be dried to a powder by any other means. Bullock and Lightbown (17) report the spray drying of digitalis, adrenalin and ascorbic acid, as well as a quantity of organic and inorganic salts.

Sewage sludge (65), distillers solubles (51, 102), latex (5, 128), and hydrocarbon catalyst particles (147) have been variously reported to have been dried by the spray process. Spray drying is used in the Dow magnesium process to concentrate the magnesium chloride solution from 15 to 35 percent. It was found that the scaling due to calcium sulphate present in the solution prevented the use of a conventional evaporator (107).

Spray drying can be used for solutions, slurries or pastes (83.) In a very general way, it may be said that if the liquor can be pumped, it can be spray dried.

V ECONOMICS OF SPRAY DRYING

The actual operating efficiencies of commercial units vary to a large extent with the type of material handled. Sherwood (108) claims that efficiencies range from 20 percent for low temperature operation to 70 per cent for higher temperature ranges. These figures are obviously approximate, but they give an indication of the results to be expected in normal operation. Fogler and Kleinschmidt (41) report that actual evaporation is 50 to 90 per cent of the theoretical.

The actual efficiency, however, would be less due to stack losses, furnace and duct radiation, etc. The above authors include performance data showing that in a 10-foot diameter chamber, from 1600 to 2100 B.t.u. are required per pound of water evaporated. Swenson data shows a heat requirement of about 1225 B.t.u. per pound of water vaporized in 10- to 18-foot diameter chambers (83). Performance data are variously reported by Burton (19) using potato mesh, Greene et al (49) producing egg powder, and by Waite (118) and Diamond (32) using laboratory size dryers. A very comprehensive report of the performance of a 4-foot diameter laboratory model has been prepared by F. H. Knelman (64).

Actual cost data naturally varies considerably from one installation to the next. The installed costs of Swenson dryers vary from a maximum of 90,000 dollars for an 18-foot model to a minimum of 30,000 for a 10-foot diameter chamber (83). Other data (41) are in agreement with these values. Sherwood (108) estimates the capital costs of installed dryers at from 10 to 25 dollars per cubic foot, the unit cost decreasing with increasing chamber diameter.

The energy requirements of spray drying are about 25 to 75 H.P. for the hot air fan, and 5 to 15 H.P. for pumping and atomizing the feed liquor (41). Sherwood (108) is in agreement with these figures, reporting an average of 50 H.P. for a 20-foot chamber.

The efficiency and economy of spray drying are claimed to compare favorably with other methods of drying (94), but it is usually considerations of feed and product properties which determine its usefulness for a specific purpose.

EXPERIMENTAL SECTION

INTRODUCTION

The experimental work has been divided into four general sections. The first deals with the effects of atomization on the temperature of the spray, for the purpose of verifying certain theoretical predictions concerning these effects (41). In the second section, a short study of the effect of feed temperature on spray drying capacity and efficiency is presented, as an extension to the work of F. H. Knelman (64). The third section, a study of the effects of feed concentration on capacity and efficiency, comprises the main body of the investigation. The last section deals with the analysis of the physical properties of the feed and of the dried product. Of particular concern in this study was the effect of feed concentration on the size of the atomized droplets and of the dried particles.

Except for a few runs in which water was the feed liquid, most of the investigations were carried out using solutions of Lignosol. This waste product of the sulphite paper industry, consisting essentially of calcium lignosulphonate, was chosen because it was used by Knelman in his previous work, and because it possesses the advantage of forming colloidal suspensions. Lignosol -- also commercially available under different trade names, such as Goulac, Bindarine, etc. -- is presently being spray dried at some Canadian mills, but very little data are available on the subject.

1. EQUIPMENT

The apparatus used in this project was originally built by Knelman and used in his investigation (64). From the general layout shown in figure 1, it can be seen that the equipment was very similar to a commercial concurrent spray dryer. It consisted essentially of a furnace to heat the drying gas, a conical chamber in which the actual spray drying took place, and a collector to recover the dried product. It was found necessary to modify certain parts of the equipment to facilitate its control and operation, but the modifications introduced were of a minor nature.

(a) Heating of the Drying Gas

The drying air was heated directly in a specially-built, gas-fired furnace, shown in figure 2. The furnace had a total volume of about 8 cu.ft., and consisted of two chambers: the combustion chamber proper, with a capacity of approximately 6 cu.ft., and the mixing chamber directly above it. The two chambers were connected by a 6 in. x 4 in. opening at the rear of the furnace. The framework of the furnace was of angle iron, the walls of 20-gage galvanized steel, and the floor and ceiling of ¼-in. steel plate. The combustion chamber was further lined with fire bricks cemented with fire clay. A damper at the back of the furnace allowed secondary air to be admitted to the mixing chamber. The furnace was operated by two gas burners, supplied from parallel gas lines, a large one for the main heat source and a smaller one for fine control. An 8-in. duct from

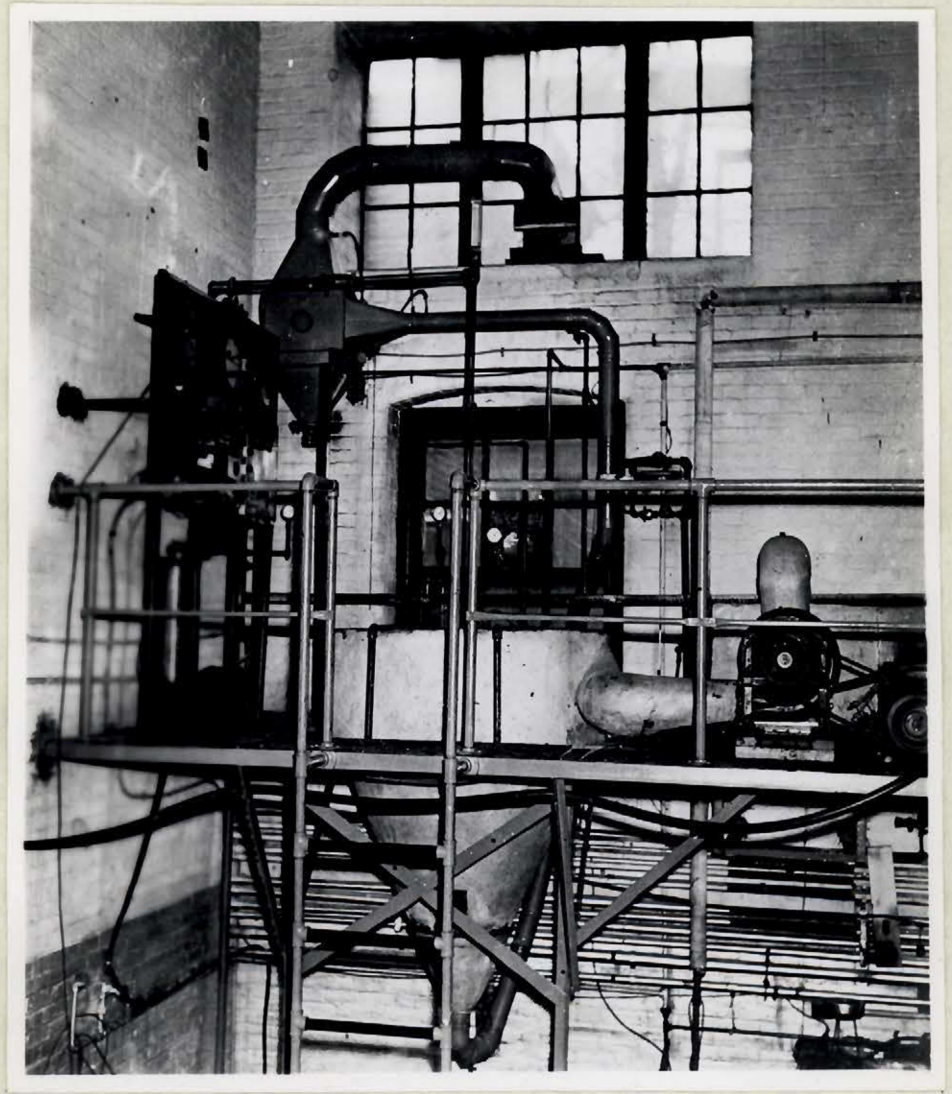

FIGURE 1

GENERAL INSTALLATION OF SPRAY DRYER
AND AUXILIARIES

FIGURE 2

FURNACE AND GAS BURNER ARRANGEMENT

the top of the mixing chamber led to the suction side of the blower. The motor-blower assembly consisted of a 3-H.P., 220-volt, D.C., compound Bogue electric motor, belt-connected to an American Blower Company, model RV-2 blower. Finally, a short length of 8-in. galvanized duct connected the blower outlet directly to the spray drying chamber. Both the furnace and the duct system were heavily insulated with Tartan high-temperature insulating cement.

(b) The Drying Chamber

A diagram, showing the chamber design, is given in figure 3. It consisted of an upper cylindrical, jacketed section, welded to a lower, conical section, the dimensions of which are indicated on the diagram. The drying gas from the blower entered the jacket tangentially and was admitted to the chamber itself through inclined slots cut out of the wall (see figure 4). The chamber was constructed entirely of 20-gage galvanized steel. The lower part of the conical section was connected to the rest of the chamber by means of flanges, and could thus be removed for cleaning purposes. As in the case of the furnace, the entire surface was covered with insulating cement. Inspection windows were installed, one in the cone and one at the top, and a light provided inside and at the top of the chamber.

Atomization of the feed was obtained by means of a two-fluid nozzle (Spraying Systems Co., model $\frac{1}{4}$ JN) using compressed air as the atomizing fluid. Both liquid feed and compressed air were supplied to the nozzle by means of $\frac{1}{4}$-in. brass lines which entered the top of the chamber through packed glands. To allow inspection and servicing of the nozzle, these glands were mounted on a circular steel plate,

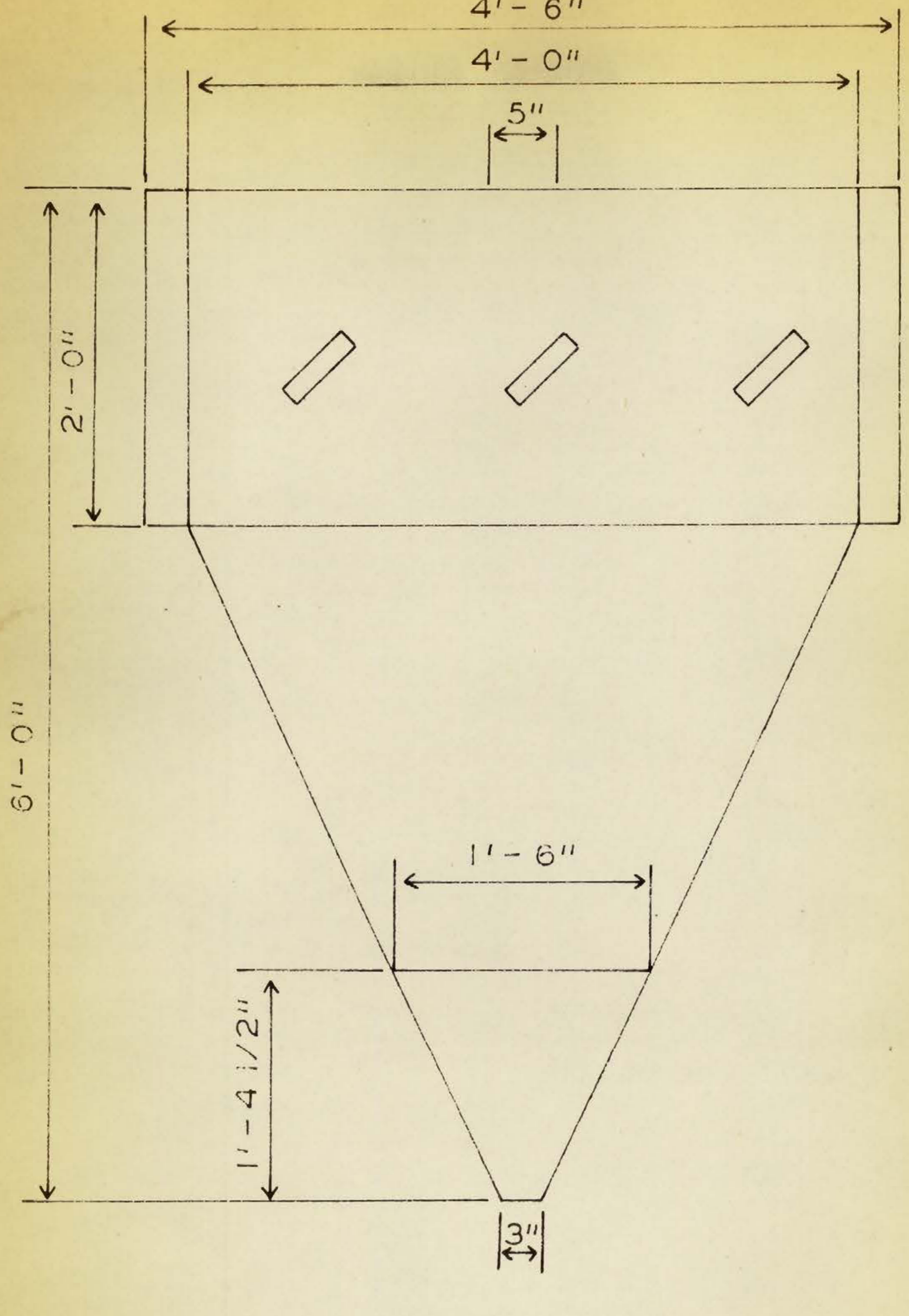

FIGURE 3

CHAMBER DESIGN

FIGURE 4

DRYING AIR INLET LOUVER

7 in. in diameter, which was in turn bolted to the top of the chamber. The nozzle assembly is illustrated in figure 5 and its position in the chamber is shown in figure 6. Short lengths of pressure tubing connected the feed and air lines to the nozzle assembly, thus allowing the depth of the latter in the chamber to be altered at will.

The solution to be dried was fed from a two-gallon stainless steel tank through a $\frac{1}{4}$-in. gear pump, and its weight rate of flow was determined by means of a calibrated rotameter.

(c) Collection of the Product

The dried Lignosol was separated from the exit gas stream in a commercial Aerotec dust collector. This apparatus consisted of twenty-five cyclone tubes arranged in rows of five, in one common sheet metal housing. The waste gases and the powder were conveyed from the bottom of the chamber to the unit through a 4-in. duct, while the dust-free gases left the collector through an 8-in. duct leading to the outside atmosphere.

(d) Control and Recording of Variables

Although adequate provision existed in the apparatus for automatic control and recording, it was found more expedient in this investigation to control all variables manually. The volumetric rate of flow of the compressed air to the nozzle was metered by means of an orifice calibrated against a wet-test meter. The feed to the nozzle was controlled with a needle-valve, and its rate of flow was measured with a calibrated rotameter. Provisions were made to determine the temperature and pressure of both streams immediately before they reached the nozzle assembly. The volumetric rate of the

FIGURE 5

SPRAY NOZZLE ASSEMBLY

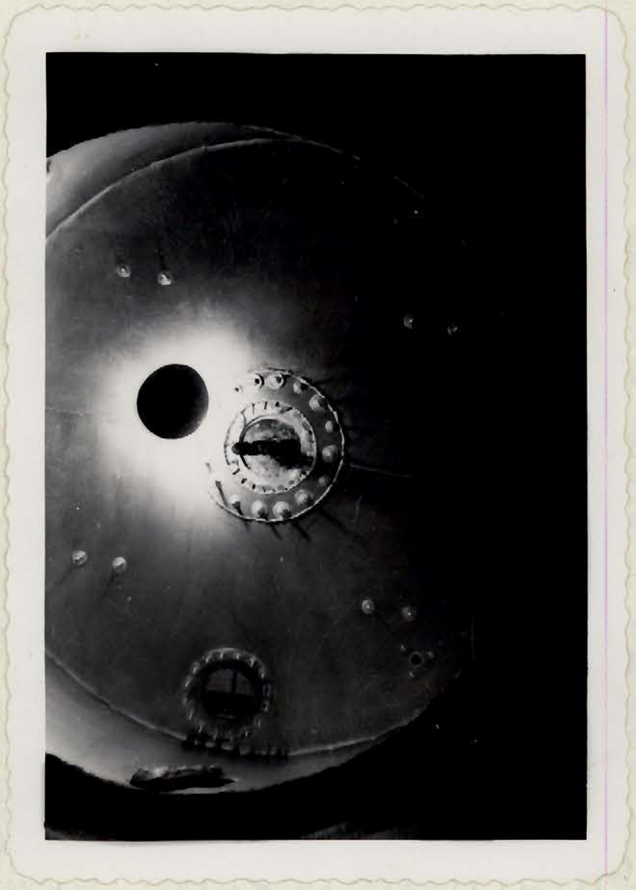

FIGURE 6

INTERIOR VIEW OF DRYING CHAMBER

hot gases to the chamber was controlled by varying the speed of the motor by means of a set of resistences inserted in the power supply to the armature. Measurement of the pressure drop through the Aerotec collector, as read on a water manometer connected across the unit, was used to determine the rate of flow of the hot gases through the chamber, from calibration charts supplied by the manufacturer.

The rate at which the gas was supplied to the furnace, and therefore the temperature of the drying gas, was also manually controlled. Provisions were made to measure the temperature and static pressure of the hot gases at various points in the system, namely: in the entering and in the exit air ducts, in the chamber jacket, and at the entrance to the Aerotec unit. In addition to these, five thermometers were imbedded at various points on the outside surface of the chamber insulation to permit calculation of the heat losses, based on data supplied by the manufacturers of the insulating material, correlating heat losses as a function of surface temperature.

The humidity of the air was determined at the chamber outlet by the well known psychrometric method. Great care was taken that the wet-bulb thermometer was introduced into the gas stream at or near the wet-bulb temperature. The humidity of the entering gases was taken at the same point before the spray was turned on, at the beginning of a trial. It was checked again by the same method at the end of the trial, but in general, negligible changes were observed in the humidity of the entering air, even over long periods of time.

2. GENERAL PROCEDURE

The same general procedure was followed for all the determinations except those concerned with the effects of atomization on spray temperature. Following adjustment of the nozzle to the required depth, the motor resistance was set to give the proper air velocity, and the motor started. With the blower thus creating a draft into the furnace, the city gas supply was turned on and ignited. The rate of flow of the fuel gas was increased very gradually until the inlet air attained the required temperature. This step generally took about half an hour.

When the proper air temperature was reached, wet- and dry-bulb temperature readings were taken at the point where the exhaust air duct joined the cone of the drying chamber. From this data, the inlet humidity of the drying gases could be calculated.

The compressed air to the nozzle was then turned on and set at the required pressure. Finally, the feed pump was started, using water as the feed during the initial stage and changing over to the Lignosol solution under investigation after a few minutes of operation. The feed rate was gradually increased until the maximum capacity of the chamber at steady state, for the given conditions, was reached. At this point, all the temperatures, pressures, flow rates and incidental data were recorded. Upon the completion of a run, water was again sprayed into the chamber to clear the feed line and the nozzle of the Lignosol test solution.

The criterion used to determine the maximum capacity of the chamber consisted of noting when particles of Lignosol started to adhere to the observation window, located in the conical section. As the chamber was flushed out after each set of runs, and this window carefully cleaned, wet particles of Lignosol adhering to the window could easily be detected. Dry particles, on the other hand, did not adhere at all, but were carried away by the gas stream.

It has been mentioned above that manual controls were used exclusively in the experimental runs. In general, only occasional adjustments were necessary to keep the operating variables constant, once steady state conditions had been attained. A source of minor irritation was the poor control of the rate of flow of the Lignosol feed solution. Because the range of velocities used was well below the operating capacity of the gear pump, the feed rate tended to decrease with time, and it was found necessary to have an operator constantly correcting for this effect.

3. CALCULATIONS AND TABLE NOMENCLATURE

The thermal efficiency of the chamber for the spray drying runs was calculated using Knelman's equation, page 38. Similarly, his definition of chamber capacity in terms of pounds of water evaporated per hour has been retained, irrespective of the nature of the feed liquid. For purposes of presentation, the experimental data presented in sections II and III have been divided into three tables: in the first are listed the observed readings; the second represents a material balance and gives the capacity; it also shows the humid heats of the drying medium; while in the third, all the terms of the heat balance leading to the calculation of the thermal efficiency are indicated.

The average droplet diameter was calculated by means of the Nukiyama and Tanasawa equation, page 10. This relationship defines the average droplet diameter, D_o, as that of a droplet with the same ratio of surface to volume as the total sum of the drops in the spray. The average particle diameter, D_p, was defined in an identical manner, thus affording direct comparison between these two physical dimensions. For the determination of D_p, the sedimentation rates of the dried Lignosol in benzene were first obtained with a Joly balance. These sedimentation rates were then used in Stokes' Law:

$$D = 20{,}000 \sqrt{\frac{9}{2}\mu \frac{dx/dt}{(\rho_p - \rho_m)980}} \qquad (13)$$

to obtain the mean particle diameter, D, in microns, for various weight limits. This equation permits calculation of the diameter, D, of a particle of density ρ_p moving with velocity dx/dt through a medium of density ρ_m and viscosity μ. In the present investigation, values of $\rho_p = 1.8$ gm./cc., $\rho_m = 0.879$ gm./cc. and $\mu = 0.00638$ poise were used. The average particle diameter, D_p, was calculated from the particle diameters obtained from Stokes' Law by means of the following relationship:

$$D_p = \frac{D^3\,dn}{D^2\,dn} \qquad (14)$$

This can be very simply integrated, either graphically or analytically, if n is taken as the weight per cent of Lignosol settling on the pan of the Joly balance. The integration limits in this case would be zero to 100 per cent.

Owing to lack of space, table headings in sections II, III and IV had to be restricted to symbols only, the full nomenclature for which is presented below:

C = Feed concentration, in weight per cent.

D_o = Average droplet diameter, in microns.

D_p = Average particle diameter, in microns.

E_D = Drying thermal efficiency, in per cent.

F = Rate of flow of feed solution, in lb./hr.

G = Weight rate of flow of drying gases, in lb./hr.

H_1 = Humidity of inlet gases, in lb./lb.

H_2 = Humidity of exit gases, in lb./lb.

l = Nozzle depth, in inches.

P_A = Nozzle air pressure, in p.s.i.g.

P_L = Feed pressure, in p.s.i.g.

Q_1 = Heat in inlet gases, in B.t.u./hr.

Q_2 = Heat in exit gases, in B.t.u./hr.

Q_F = Heat in feed solution, in B.t.u./hr.

Q_L = Heat losses, in B.t.u./hr.

Q_V = Heat used to vaporize the water in the feed, in B.t.u./hr.

Q_a = Volumetric rate of flow of atomizing air, in c.f.m.

Q_l = Volumetric rate of flow of feed liquid, in c.f.m.

S_1 = Humid heat of inlet gases, in B.t.u./(hr.)(°F.)

S_2 = Humid heat of exit gases, in B.t.u./(hr.)(°F.)

T_1 = Inlet drying gas temperature, in °F.

T_2 = Exit drying gas temperature, in °F.

T_A = Atomizing air temperature, in °F.

T_F = Inlet liquid feed temperature, in °F.

T_{WB} = Wet-bulb temperature of drying gases, in °F.

V = Volumetric rate of flow of drying gases, in c.f.m.

v = Relative velocity of atomizing air with respect to feed liquid, $= v_a - v_l$, in m./sec.

v_a = Velocity of atomizing air at the throat of the nozzle, in m./sec.

v_l = Velocity of feed liquid at the throat of the nozzle, in m./sec.

W = Chamber capacity, in lb. of water evaporated/hr.

λ_{WB} = Latent heat of vaporization of water at the wet-bulb temperature, in B.t.u./lb.

ρ = Density, in gm./cc.

σ = Surface tension, in dynes/cm.

μ = Viscosity, in poises.

I. EFFECT OF ATOMIZATION ON SPRAY TEMPERATURE

In 1938, Fogler and Kleinschmidt (41) predicted that, owing to adiabatic expansion, large temperature drops should be observed through two-fluid nozzles -- such as were used in this investigation -- during atomization. Such pronounced cooling effects in the spray would, in turn, exert a significant influence on the subsequent rate of drying in the chamber, and it was for the purpose of ascertaining their magnitude that this section of the experimental work was undertaken. The Nukiyama and Tanasawa equation (page 10) shows that the degree of atomization is essentially a function of the velocity of the compressed air stream relative to that of the liquid feed, and of the physical properties of the latter. Since these, in turn, are functions of the liquid and compressed air temperatures and pressures, the latter were varied over a wide range of operating conditions and the temperature of the spray measured simultaneously.

1. EQUIPMENT AND PROCEDURE

For this set of runs, the nozzle was removed from the chamber and set up to spray directly into the atmosphere. The cold junction of a copper-constantan thermocouple was so arranged that it could be held in the spray at varying distances from the nozzle outlet, while the hot junction was immersed in water boiling at atmospheric pressure. A Leeds and Northrup potentiometer, model No. 8657C, was used to measure the potential difference, and the corresponding temperature difference was obtained from calibration charts previously established on mixtures of brine and ice.

2. EXPERIMENTAL RESULTS

(a) Variation of Spray Temperature with Distance from Nozzle

Using compressed air alone at a constant pressure of 45 p.s.i.g., the temperature drop across the nozzle was measured, at varying distances from the outlet. Water was then pumped to the nozzle at a constant rate of 45 lb./hr., and similar measurements were repeated on the wet spray. The results are shown in table I and the temperature drops are plotted on figure 7.

In the case of air alone, the temperature drop reaches a maximum at about 1/8 in. from the outlet, the two opposing mechanisms of cooling due to expansion and heat transfer from the surroundings are equally balanced, while at greater distances, the rate of heat transfer to the expanding jet of gas predominates. It is significant that these temperature effects are relatively small.

When water is atomized, however, figure 7 clearly shows that the temperature drop is much greater. Under these conditions, the variation with distance from nozzle outlet is small. These rather unexpected results indicate that the cooling effect on the spray could not possibly be due to expansion of the compressed air alone to more than a slight extent, but is mainly due to evaporation of the sprayed droplets. These cooling effects, following a rapid increase during the initial travel of the droplet, should thereafter

Table I

VARIATION OF TEMPERATURE DROP WITH DISTANCE

Room Temperature	70°F.
Feed Temperature	72°F.
Feed Rate	45 lb./hr.
Air Pressure	45 p.s.i.g.

Distance from Nozzle	Temperature Drop	
	Air	Air and Water
inches	°F	°F
0	5.2	-
1/8	7.3	18
1/4	7.2	-
1	5.6	19
2	2.4	20

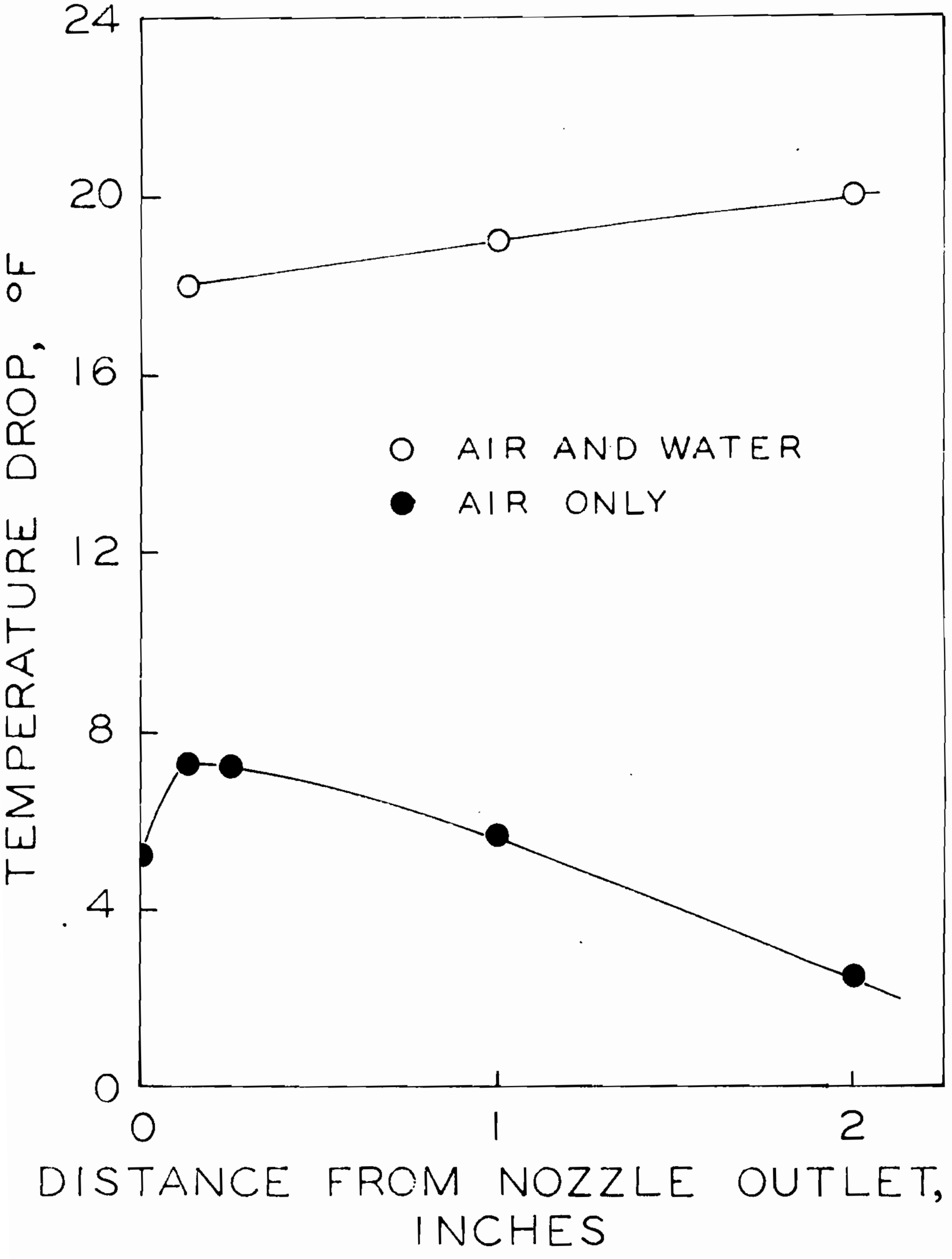

FIGURE 7

VARIATION OF TEMPERATURE DROP WITH DISTANCE FROM NOZZLE increase more slowly as the wet-bulb temperature of the air is approached, as confirmed in figure 7. It should be noted that in these trials, as well as in the subsequent parts (b) and (c), variations in the feed water rate introduced negligible differences in the results.

(b) Variation of Spray Temperature with Pressure of Atomizing Air

The effects of atomization on the temperature of the spray were further investigated by varying the compressed air pressure over a wide range, at three different feed water temperatures. The thermocouple junction was placed at a fixed distance of 1/8 in. from the nozzle, and the water rate was maintained constant at 75 pounds per hour during these trials. The results are shown in table II and on figure 8. For convenience actual spray temperatures, instead of temperature drops, are shown on the latter.

For all three initial feed temperatures, an increase in atomizing air pressure causes a decrease in the temperature of the spray. Since an increase in pressure causes a decrease in drop size and hence an increased rate of evaporation, (64), these results appear to lend additional support for the mechanism of evaporation as the main factor in the cooling of a spray during atomization. However, the wet-bulb temperature of the air constitutes the practical lower limit below which no further cooling by evaporation can take place. Under the actual operating conditions, the wet-bulb temperature of the air was 65°F., and the curves of figure 8 clearly indicate convergence towards this common lower temperature. In the light of the results shown in figure 7, there is little doubt that

Table II

VARIATION OF SPRAY TEMPERATURE WITH ATOMIZING AIR PRESSURE

Room Temperature	77°F.
Wet-Bulb Temperature	65°F.
Humidity	0.0105 lb./lb.
Feed Rate	75 lb./hr.
Distance from Nozzle	1/8 in.

Air Pressure	Inlet Feed Temperature, °F.		
	81	107.5	122
p.s.i.g.	Spray Temperature, °F.		
0	78.4	104.9	118.2
10	70.5	80.6	92.0
20	68.0	74.3	83.8
30	68.0	70.0	80.0
40	65.4	69.8	75.8
45	65.4	69.8	75.0

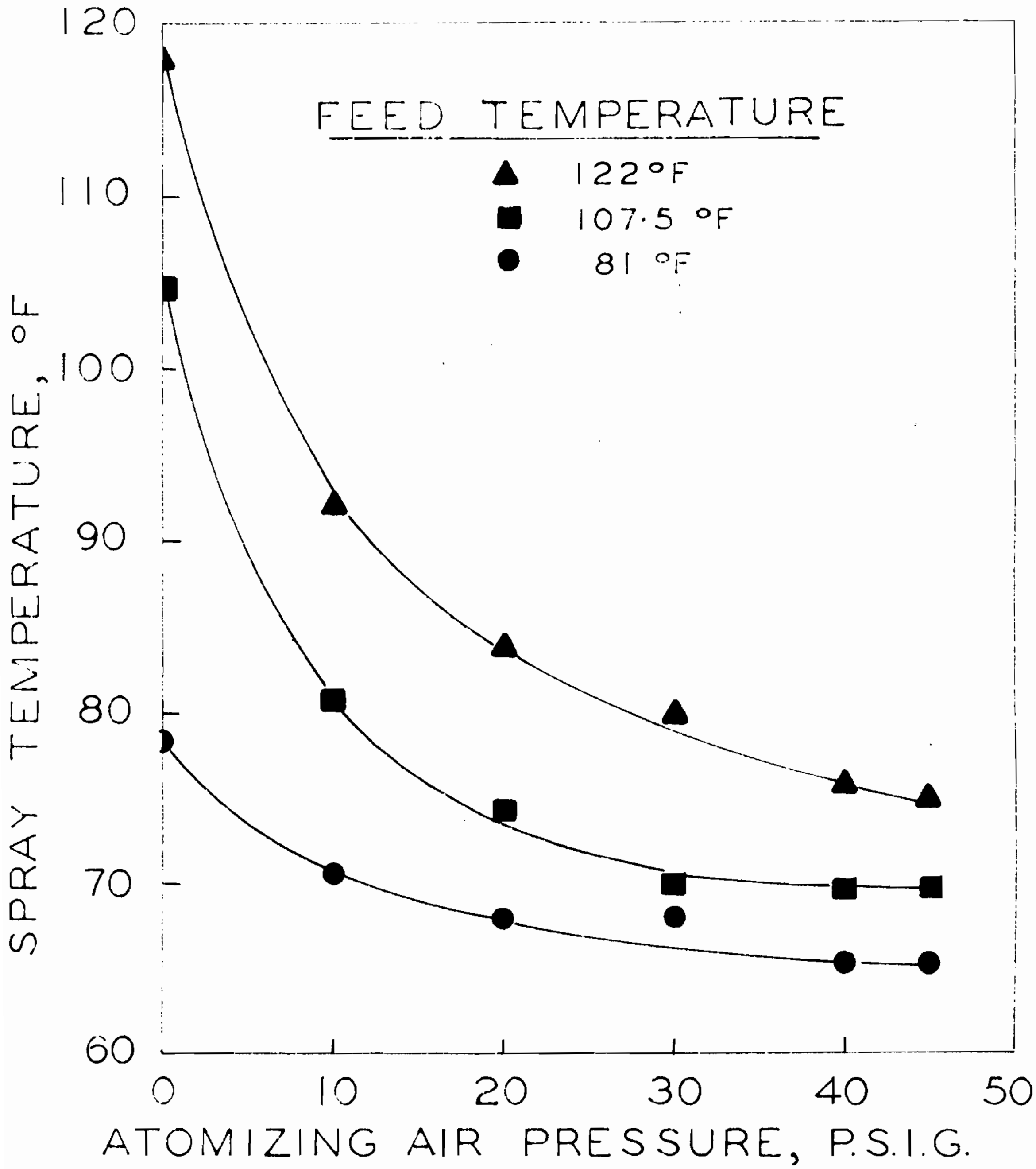

FIGURE 8

VARIATION OF SPRAY TEMPERATURE WITH ATOMIZING AIR PRESSURE this convergence would have been more sharply indicated, had the temperature of the spray been measured at a greater distance from the nozzle outlet. It should be noted that when no compressed air was used, the passage of feed water alone through the nozzle was accompanied by a temperature drop of only a few degrees Fahrenheit (figure 8, zero air pressure). This slight cooling effect should probably be ascribed to heat transfer to the nozzle assembly by conduction, as no atomization is obtained under these conditions.

(c) Variation of Spray Temperature with Inlet Feed Temperature

The final set of runs of this series was made to determine in more detail than was done in part (b) the relationship existing between spray temperature and feed temperature. The feed was kept below its normal boiling point to avoid flashing effects. The feed rate and compressed air pressure were maintained constant at 45 pounds per hour and 40 p.s.i.g., respectively, and the cold junction located at a fixed distance of 1/8 in. from the nozzle. The room temperature and humidity for this entire set of runs were 77°F., and 0.011 lb./lb. The results are shown in table III and on figure 9.

The temperature of the spray is seen to rise only slightly as the feed temperature is increased. In the light of the previous explanation, cooling effects due to evaporation should lead, at steady state, to a spray temperature corresponding to the wet-bulb temperature of the air, namely: 65°F., in this set of runs. The values shown on figure 9 indicate that steady state was not completely attained at the particular location of the cold junction selected, and that departure from equilibrium conditions was more pronounced as the temperature of the feed was increased.

Table III

VARIATION OF SPRAY TEMPERATURE WITH FEED TEMPERATURE

Room Temperature	77°F.
Wet-Bulb Temperature	66°F.
Humidity	0.011 lb./lb.
Feed Rate	45 lb./hr.
Air Pressure	40 p.s.i.g.
Distance from Nozzle	1/8 in.

Feed Temperature °F.	Spray Temperature °F.
83.3	65.7
102.2	65.7
113.9	68.0
139.1	75.6
156.2	78.0
164.3	78.0

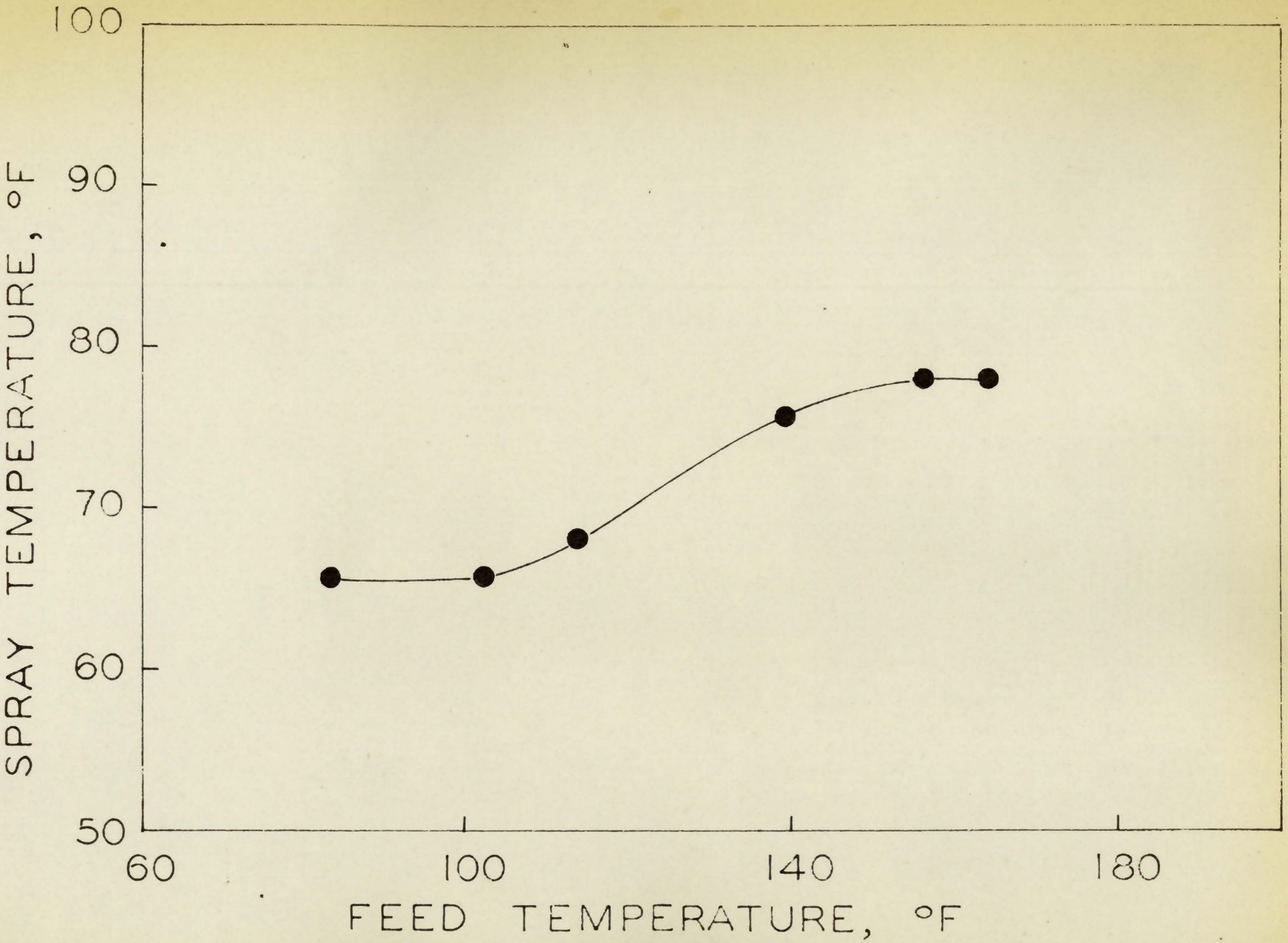

FIGURE 9

VARIATION OF SPRAY TEMPERATURE
WITH FEED TEMPERATURE

II. EFFECT OF FEED TEMPERATURE ON CHAMBER CAPACITY AND DRYING EFFICIENCY

In a previous study of the effects of feed temperature on the spray drying processes (64), Knelman had shown that the capacity of the chamber passed through a maximum when the feed -- whether water or a Lignosol solution -- entered at the wet-bulb temperature of the drying gases. However, in the light of the work done in the previous section, it was felt that not enough determinations had been made above the wet-bulb temperature to justify this conclusion. The present set of runs was made to obtain the missing data.

1. EQUIPMENT AND PROCEDURE

The equipment and procedure for determining the capacity and efficiency were exactly as outlined in the introductory section, pages 44 et seq. Pure water was used as feed in this entire set of runs, and its temperature was maintained by means of a steam-operated heater installed in the feed tank.

2. EXPERIMENTAL RESULTS

The operating conditions selected were approximately the same as those used by Knelman, namely: a drying gas rate of 262 c.f.m. (free air), a nozzle depth of 11.25 inches, and an inlet drying gas temperature of 420°F. The results are shown in table IV and on figure 10.

WATER-VARIATION OF INLET FEED TEMPERATURE

Table IV(a)

Trial No.	l in.	V cfm	T_1 °F	T_F °F	T_2 °F	T_{WB} °F	P_A psig.	P_L psig.
1	11.25	262	420	86.9	149	122	40	40
2			424	98.6	151			
3			415	113.9	142			
4			423	131.0	144			
5			418	143.6	147			
6			423	152.6	153			
7			420	170.6	149			

Table IV(b)

Trial No.	F lb/hr	H_1 lb/lb	H_2 lb/lb	S_1 Btu/lb	S_2 Btu/lb	G lb/hr	W lb/hr
1	49.8	0.0285	0.0712	0.253	0.272	1170	49.8
2	53.0		.0740		.274		53.0
3	53.7		.0748		.274		53.7
4	55.8		.0760		.274		55.8
5	57.4		.0774		.275		57.4
6	57.4		.0776		.275		57.4
7	57.4		.0775		.275		57.4

Table IV(c)

Trial No.	λ W.B. Btu/lb	Q_1 Btu/hr	Q_2 Btu/hr	Q_V Btu/hr	Q_F Btu/hr	Q_L Btu/hr	E_D %
1	1024	88,200	8,900	51,000	-1,750	26,550	59.0
2		89,500	9,300	54,400	1,240	24,560	61.5
3		86,600	6,405	55,000	435	24,760	63.8
4		89,000	7,050	57,100	+ 502	25,352	64.0
5		87,500	8,045	58,600	1,240	22,095	66.0
6		89,000	9,955	58,600	1,760	22,205	64.6
7		88,200	8,690	58,600	2,800	23,710	64.5

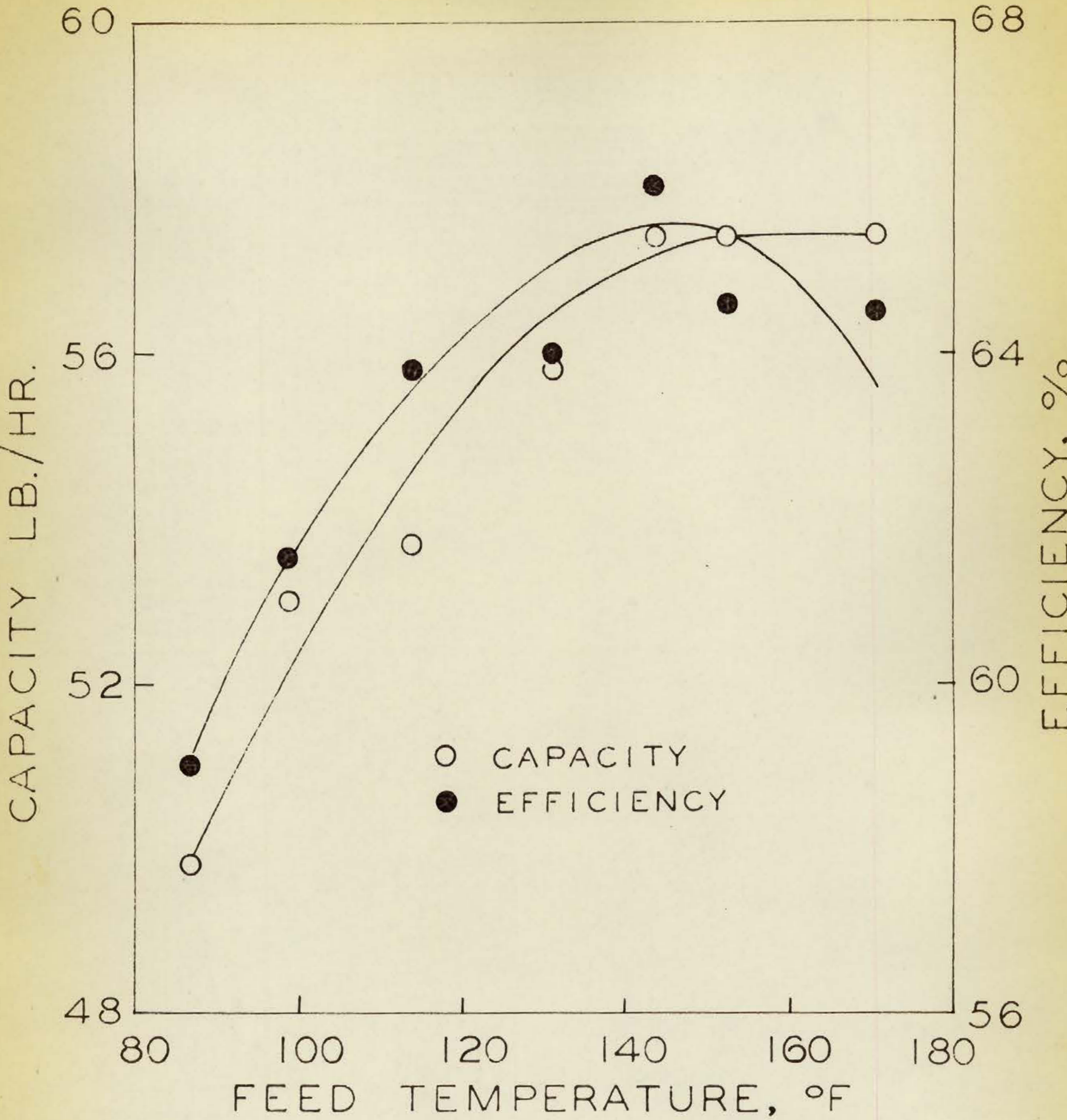

FIGURE 10

VARIATION OF CAPACITY AND EFFICIENCY
WITH INLET FEED TEMPERATURE

The curves show that, contrarily to Knelman's work, the capacity curve does not pass through a maximum, but tends to approach a constant value at high feed temperatures. The efficiency curve, however, indicates a maximum, and the values reported are in good agreement with Knelman's results.

III. EFFECT OF FEED CONCENTRATION ON SPRAY DRYING CAPACITY AND EFFICIENCY

In his study on spray drying, Knelman made an extensive investigation of the effects of the various operating variables. However, only pure water and a Lignosol solution of fixed concentration were used. Since changes in feed composition will affect both the degree of atomization and the rate of drying, it was felt that a study of their effects on the spray drying capacity and efficiency was indicated.

1. EQUIPMENT AND PROCEDURE

The equipment and general procedure used were as outlined in the preliminary description of the experimental section, pages 44 et seq. A large number of runs had to be made to determine, at each feed concentration, the effect on capacity and efficiency of varying the nozzle depth and the drying gas temperature and volumetric rate. Four different Lignosol solutions were used, in addition to pure water. At each concentration three sets of runs were made, at three different air rates, to determine the optimum nozzle position. With the nozzle set at this position, further trials were made at a constant drying gas rate, to determine the effect of varying the temperature of the gases.

2. EXPERIMENTAL RESULTS

(a) Variation of Nozzle Position and Inlet Gas Rate

Knelman reported that the capacity of a spray drying chamber passed through a sharp maximum as the depth of the nozzle in the chamber was increased, other variables being constant, and that, further, this optimum nozzle location was markedly lower for a Lignosol solution than for pure water (64). This interesting effect was investigated at some length in the present study.

The results are set forth in tables V - IX and on figures 11 - 15. It will be noted that there is a definite optimum nozzle depth for every set of conditions. This appears to increase with increasing concentrations, but is not influenced by the volumetric rate of drying gases. The efficiency curves closely follow the same behaviour, as should be expected, since, the heat input to the chamber being constant for a given air rate, the efficiency expression becomes a direct function of the amount of water evaporated per hour, that is, of the capacity. It should be noted, however, that for any feed concentration, and for a given nozzle depth, the volumetric rate of flow leading to maximum capacity does not necessarily result in maximum efficiency. Thus, for a concentration of 5%, maximum capacity is obtained at 308 c.f.m., while maximum efficiency occurs at the lower rate of 208 c.f.m. (free air), figure 12.

The significant effect exerted by the feed concentration on the optimum location of the nozzle is best illustrated by figure 16, based on figures 11 - 15.

WATER-VARIATION OF NOZZLE DEPTH AND DRYING GAS VELOCITY

Table V(a)

Trial No.	1 in.	V cfm	T_1 °F	T_F °F	T_2 °F	T_{WB} °F	P_A psig.	P_L psig.
8	9	308	399	78.8	174	119.3	35	38
9	10		397	79.7	171		35	39
10	10.5		397	80.0	154		36	40
11	11		397	79.7	149		37	40
12	11.5		390	78.8	151		37	41
13	12		394	78.8	151		38	41
14	10	262	388	72.5	131	117.5	36	40
15	10.5		392	71.6	135			
16	11		387	68.0	131			
17	11.5		390	68.0	131			
18	12		385	69.8	135			
19	10	208	372	78.8	131	118.4	36	40
20	10.5		381	78.8	135		36	
21	11		385	81.5	124		36	
22	11.5		378	85.1	127		35	
23	12		383	85.1	133		36	

WATER-VARIATION OF NOZZLE DEPTH AND DRYING GAS VELOCITY

Table V(b)

Trial No.	F lb/hr	H_1 lb/lb	H_2 lb/lb	S_1 Btu/lb	S_2 Btu/lb	G lb/hr	W lb/hr
8	47.9	0.023	0.0575	0.250	0.266	1380	47.9
9	49.6		.0588		.267		49.6
10	60.0		.0667		.270		60.0
11	61.3		.0672		.270		61.3
12	59.0		.0659		.270		59.0
13	58.0		.0648		.269		58.0
14	59.6	0.0172	0.0680	0.248	0.271	1170	59.6
15	59.6		.0692		.271		59.6
16	61.5		.0700		.272		61.5
17	58.5		.0669		.270		58.5
18	56.0		.0655		.270		56.0
19	54.0	0.0208	0.0754	0.249	0.274	980	54.0
20	55.2		.0769		.275		55.2
21	57.7		.0791		.276		57.7
22	54.5		.0758		.274		54.5
23	51.5		.0730		.273		51.5

WATER-VARIATION OF NOZZLE DEPTH AND DRYING GAS VELOCITY

Table V(c)

Trial No.	$\lambda_{W.B.}$ Btu/lb	Q_1 Btu/hr	Q_2 Btu/hr	Q_V Btu/hr	Q_F Btu/hr	Q_l Btu/hr	E_D %
8	1026	96,500	20,400	49,100	-1,940	25,260	52.0
9		95,900	19,000	50,900	2,015	23,985	54.2
10		95,900	13,140	61,500	2,430	18,830	65.8
11		95,900	11,270	62,900	2,480	19,250	67.4
12		93,500	12,000	60,500	2,390	18,610	66.4
13		94,900	12,000	59,500	2,350	21,050	64.1
14	1027	78,600	4,450	61,250	-2,720	10,180	80.8
15		79,800	5,400	61,250	2,740	10,410	79.5
16		78,000	4,450	63,000	3,040	7,510	84.0
17		79,200	4,450	60,000	2,900	11,850	78.6
18		77,500	5,400	57,500	2,680	11,920	77.0
19	1026	62,000	3,480	55,400	-2,140	980	92.4
20		64,200	4,280	56,500	2,185	1,435	91.1
21		65,000	1,340	59,100	2,130	2,430	94.2
22		63,200	2,400	56,000	1,820	2,980	91.5
23		64,600	3,740	52,800	1,720	6,340	84.0

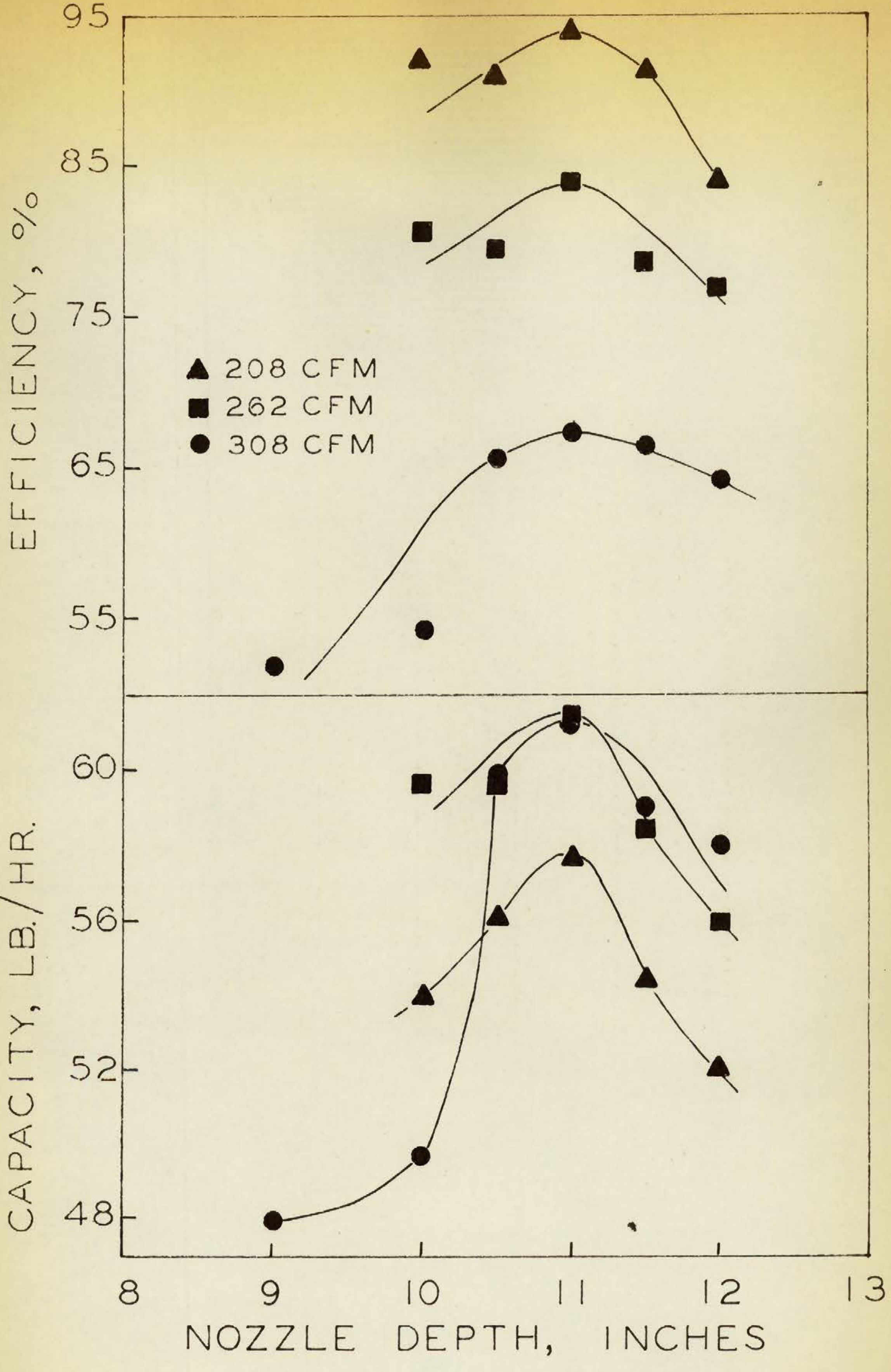

FIGURE 11

WATER - VARIATION OF CAPACITY AND EFFICIENCY WITH NOZZLE DEPTH

5% LIGNOSOL-VARIATION OF NOZZLE DEPTH AND DRYING GAS VELOCITY

Table VI(a)

Trial No.	l in.	V cfm	T_1 °F	T_F °F	T_2 °F	T_{WB} °F	P_A psig.	P_L psig.
28	9	308	396	87.8	282	122.0	36	35
29	10		392	89.6	277			
30	11.5		394	94.1	275			
31	12		401	95.0	282			
32	13		406	93.2	284			
33	9	262	385	95.0	266	116.6	35	32
34	10		387	93.2	261		36	
35	11.5		383	95.0	259		36	
36	12		387	95.0	266		36	
37	13		388	95.0	268		36	
38	9	208	379	100.4	241	116.6	36	36
39	10		381	98.6	234			
40	11.5		383	97.7	230			
41	12		381	97.5	246			
42	13		381	95.0	246			

5% LIGNOSOL-VARIATION OF NOZZLE DEPTH AND DRYING GAS VELOCITY

Table VI(b)

Trial No.	F lb/hr	H_1 lb/lb	H_2 lb/lb	S_1 Btu/lb	S_2 Btu/lb	G lb/hr	W lb/hr
28	19.3	0.026	0.0396	0.252	0.258	1380	18.37
29	20.9		.0406		.258		19.82
30	21.3		.0410		.259		20.20
31	20.9		.0400		.258		19.82
32	18.7		.0390		.258		17.80
33	18.0	0.017	0.0320	0.248	0.254	1170	17.10
34	20.2		.0332		.255		19.29
35	20.8		.0342		.256		19.82
36	18.5		.0321		.254		17.60
37	18.5		.0320		.254		17.60
38	18.0	0.012	0.0295	0.245	0.253	980	17.10
39	19.3		.0310		.254		18.30
40	20.3		.0317		.254		19.29
41	17.5		.0290		.253		16.60
42	16.7		.0281		.253		15.90

5% LIGNOSOL-VARIATION OF NOZZLE DEPTH AND DRYING GAS VELOCITY

Table VI(c)

Trial No.	$\lambda_{W.B.}$ Btu/lb	Q_1 Btu/hr	Q_2 Btu/hr	Q_V Btu/hr	Q_F Btu/hr	Q_L Btu/hr	E_D %
28	1024	95,300	57,600	18,800	-628	18,272	19.85
29		94,000	55,800	20,350	642	17,208	21.80
30		94,500	55,000	20,700	565	18,235	22.00
31		97,000	57,600	20,350	535	18,515	21.05
32		98,800	58,400	18,240	514	21,646	18.58
33	1027	77,800	44,500	17,550	-370	15,380	22.70
34		78,400	43,000	19,800	416	15,184	25.40
35		77,100	42,500	20,400	429	13,771	26.60
36		78,400	44,500	18,090	380	15,430	23.20
37		78,900	45,100	18,090	380	15,330	23.00
38	1027	63,100	31,400	17,540	-277	13,883	28.00
39		63,600	29,600	18,800	330	14,870	29.75
40		64,000	28,600	19,790	364	15,246	31.00
41		63,600	33,000	17,020	328	13,252	26.90
42		63,600	33,000	16,300	344	13,956	25.80

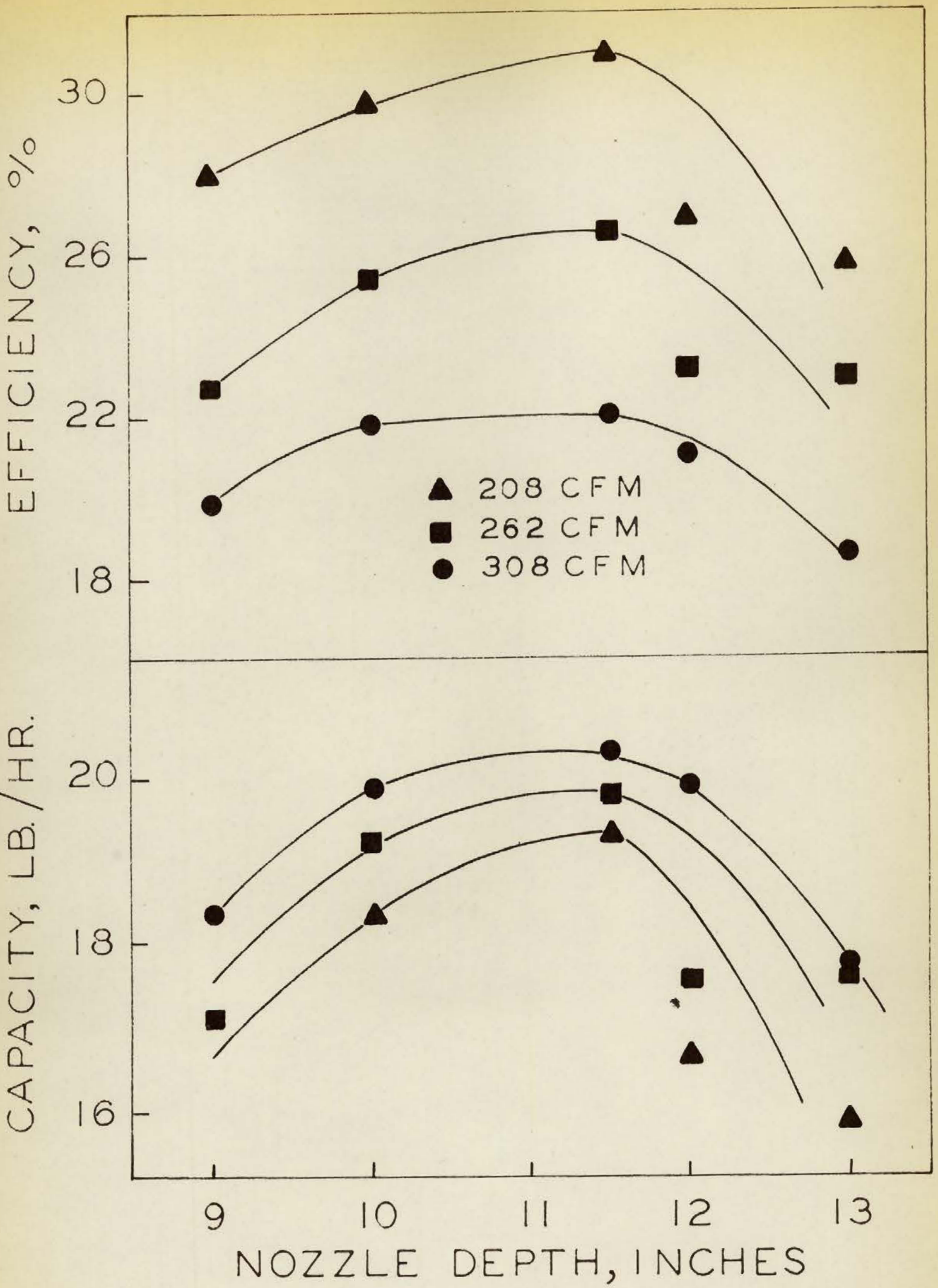

FIGURE 12

5% LIGNOSOL - VARIATION OF CAPACITY AND EFFICIENCY WITH NOZZLE DEPTH

7½% LIGNOSOL-VARIATION OF NOZZLE DEPTH AND DRYING GAS VELOCITY

Table VII(a)

Trial No.	l in.	V cfm	T_1 °F	T_F °F	T_2 °F	T_{WB} °F	P_A psig.	P_L psig.
47	10	308	401	95.0	293	122.0	36	29
48	11		401	95.0	293		37	30
49	12		397	90.5	282		38	30
50	13		406	95.9	300		36	30
51	10	262	392	88.7	261	118.4	35	30
52	11		385	91.4	250			29
53	11.5		385	95.0	250			30
54	12		379	95.9	253			31
55	12.5		387	93.2	255			31
56	13		378	94.1	255			31
57	11	308	387	86.0	244	117.5	36	31
58	11.5		394	86.0	244			31
59	12		396	86.0	248			31
60	12.5		397	86.0	255			30
61	13		401	84.2	259			30

7½% LIGNOSOL-VARIATION OF NOZZLE DEPTH AND DRYING GAS VELOCITY

Table VII(b)

Trial No.	F lb/hr	H_1 lb/lb	H_2 lb/lb	S_1 Btu/lb	S_2 Btu/lb	G lb/hr	W lb/hr
47	15.6	0.0330	0.0432	0.255	0.260	1380	14.42
48	17.0		.0443		.260		15.70
49	18.4		.0454		.261		17.02
50	14.4		.0425		.259		13.31
51	17.5	0.0230	0.0368	0.250	0.257	1170	16.20
52	18.4		.0380		.257		17.02
53	18.4		.0380		.257		17.02
54	19.8		.0390		.258		18.31
55	19.2		.0384		.258		17.76
56	17.5		.0368		.257		16.20
57	17.0	0.020	0.0362	0.249	0.256	980	15.70
58	17.0		.0363		.256		15.70
59	17.9		.0370		.257		16.53
60	17.0		.0362		.256		15.70
61	15.6		.0350		.256		14.42

7½% LIGNOSOL-VARIATION OF NOZZLE DEPTH AND DRYING GAS VELOCITY

Table VII(c)

Trial No.	λ W.B. Btu/lb	Q_1 Btu/hr	Q_2 Btu/hr	Q_V Btu/hr	Q_F Btu/hr	Q_L Btu/hr	E_D %
47	1024	98,100	61,100	14,800	-390	15,810	15.14
48		98,100	61,100	16,100	424	20,476	16.49
49		96,800	57,100	17,480	536	21,684	18.13
50		100,000	63,600	13,660	348	22,392	13.70
51	1026	80,000	43,000	16,600	-481	19,919	20.90
52		77,900	39,700	17,480	460	20,260	22.60
53		77,900	39,700	17,480	398	20,322	22.55
54		76,400	41,000	18,800	413	16,187	24.80
55		78,400	41,500	18,200	446	18,254	23.35
56		75,700	41,500	16,600	394	17,206	22.00
57	1026	65,500	32,200	16,100	-495	16,705	24.80
58		67,400	32,200	16,100	495	18,605	24.10
59		68,000	33,200	17,000	520	17,280	25.20
60		68,200	34,900	16,100	495	16,705	23.80
61		69,200	35,500	14,800	480	19,020	21.55

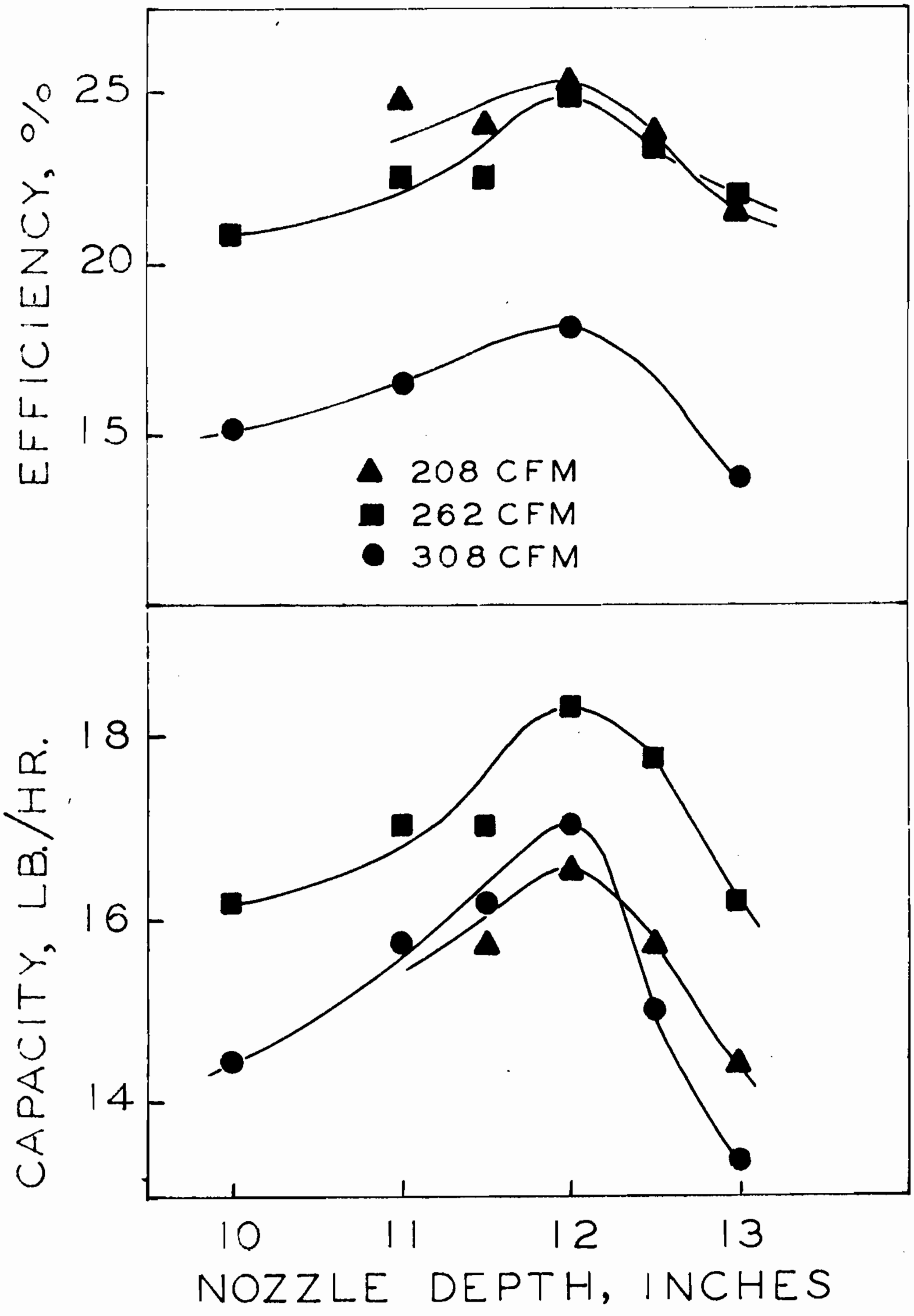

FIGURE 13

7½% LIGNOSOL - VARIATION OF CAPACITY AND EFFICIENCY WITH NOZZLE DEPTH

10% LIGNOSOL-VARIATION OF NOZZLE DEPTH AND DRYING GAS VELOCITY

Table VIII(a)

Trial No.	l in.	V cfm	T_1 °F	T_F °F	T_2 °F	T_{WB} °F	P_A psig.	P_L psig.
66	10	308	424	82.4	322	120.2	35	35
67	11		424	88.7	318			
68	12		412	91.4	311			
69	12.5		419	95.0	306			
70	13		417	90.5	313			
71	14		419	92.3	315			
72	10	262	421	93.2	298	118.4	35	35
73	11		426	95.0	306			
74	12		403	90.5	284			
75	12.5		417	93.2	282			
76	13		423	93.2	294			
77	14		430	95.0	302			
78	10	208	418	91.4	285	116.6	35	32
79	11		414	91.4	286			
80	12		415	89.6	280			
81	12.5		419	90.5	279			
82	13		410	90.5	290			
83	14		413	91.4	286			

10% LIGNOSOL-VARIATION OF NOZZLE DEPTH AND DRYING GAS VELOCITY

Table VIII(b)

Trial No.	F lb/hr	H_1 lb/lb	H_2 lb/lb	S_1 Btu/lb	S_2 Btu/lb	G lb/hr	W lb/hr
66	11.6	0.0260	0.0334	0.252	0.255	1380	10.40
67	12.9		.0343		.256		11.60
68	14.9		.0354		.256		13.40
69	17.0		.0370		.257		15.32
70	14.3		.0353		.256		12.90
71	13.5		.0346		.256		12.10
72	14.3	0.0215	0.0325	0.250	0.255	1170	12.91
73	15.6		.0335		.255		14.05
74	17.0		.0344		.256		15.60
75	17.9		.0354		.256		16.10
76	17.0		.0347		.256		15.60
77	17.0		.0346		.256		15.60
78	11.6	0.0240	0.0343	0.251	0.256	980	10.42
79	12.1		.0350		.256		10.86
80	13.7		.0366		.256		12.39
81	15.1		.0379		.257		13.58
82	12.4		.0350		.256		11.12
83	11.9		.0349		.256		10.75

10% LIGNOSOL-VARIATION OF NOZZLE DEPTH AND DRYING GAS VELOCITY

Table VIII(c)

Trial No.	λ W.B. Btu/lb	Q_1 Btu/hr	Q_2 Btu/hr	Q_V Btu/hr	Q_F Btu/hr	Q_L Btu/hr	E_D %
66	1025	105,900	70,800	10,670	-437	23,993	10.10
67		105,900	69,500	11,900	394	24,106	11.28
68		101,500	67,000	13,720	429	20,351	13.59
69		104,000	65,300	15,700	429	22,571	15.15
70		103,100	67,700	13,210	413	21,777	12.90
71		104,000	68,500	12,400	364	22,736	11.99
72	1026	88,400	53,500	13,250	-362	21,288	15.09
73		90,000	55,400	14,400	365	19,835	16.08
74		83,000	49,500	16,000	484	17,016	19.41
75		87,500	49,000	16,500	450	21,550	18.96
76		89,000	52,400	16,000	437	20,163	18.08
77		91,000	54,500	16,000	416	20,084	17.62
78	1028	74,300	44,100	10,710	-263	19,227	14.49
79		73,000	41,900	11,140	274	19,686	15.31
80		73,500	40,600	12,700	334	19,866	17.35
81		74,300	40,100	13,920	354	19,926	19.11
82		72,100	43,100	11,420	290	17,290	15.92
83		72,900	41,900	11,030	271	19,699	15.20

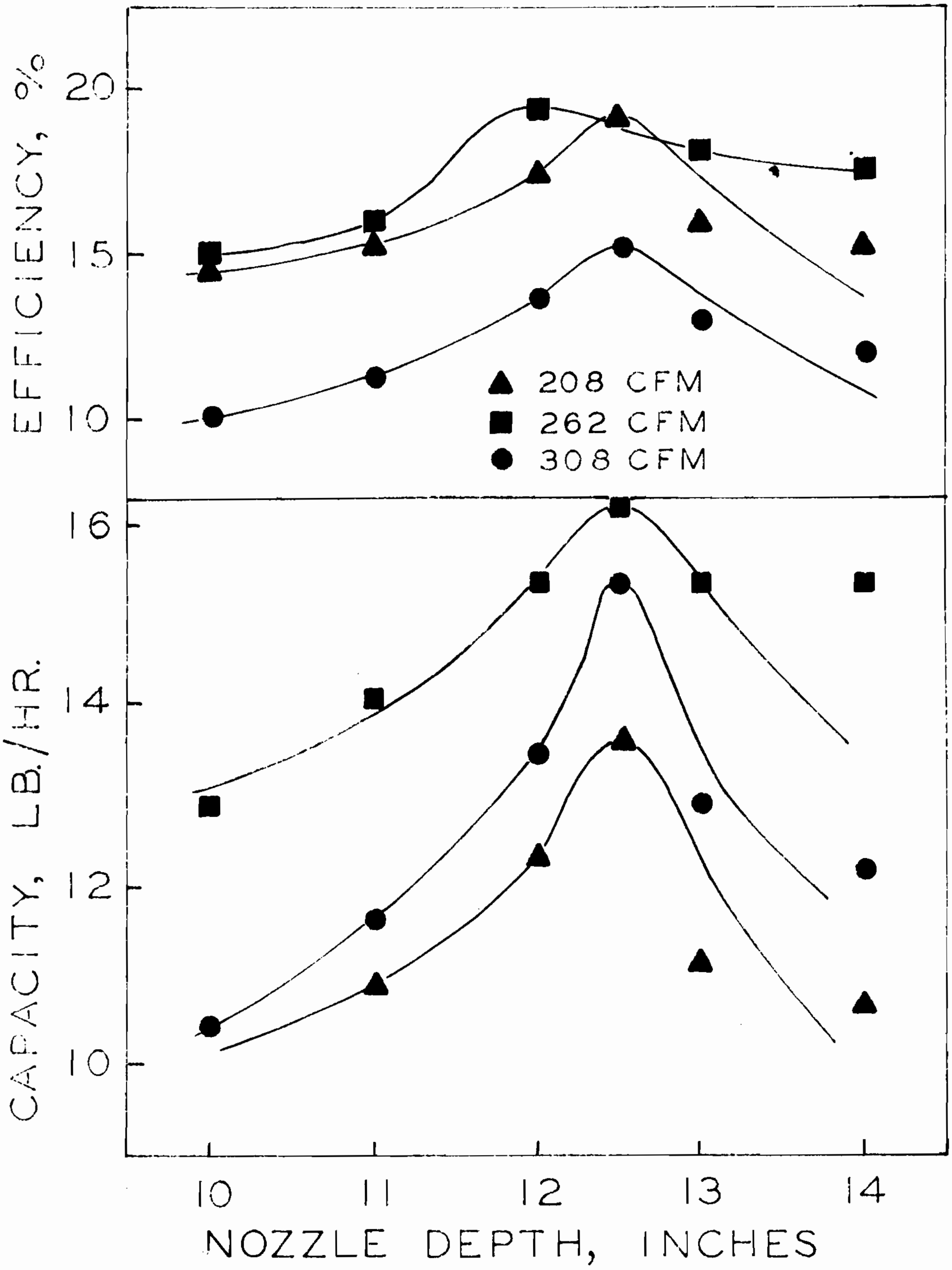

FIGURE 14

10% LIGNOSOL - VARIATION OF CAPACITY AND EFFICIENCY WITH NOZZLE DEPTH

17½% LIGNOSOL-VARIATION OF NOZZLE DEPTH AND DRYING GAS VELOCITY

Table IX(a)

Trial No.	l in.	V cfm	T_1 °F	T_F °F	T_2 °F	T_{WB} °F	P_A psig.	P_L psig.
89	12	308	396	85.1	325	116.6	42	48
90	13		390	85.1	320		42	47
91	13.5		393	86.0	322		45	47
92	14		390	86.9	327		44	46
93	11	262	390	87.8	318	116.6	50	44
94	12.5		394	87.8	315		51	
95	13		392	89.6	316		46	
96	13.5		392	91.4	322		43	
97	11	208	392	93.2	315	117.5	48	44
98	12.5		397	93.2	315			
99	13		397	93.2	327			
100	13.5		390	93.2	327			

17½% LIGNOSOL-VARIATION OF NOZZLE DEPTH AND DRYING GAS VELOCITY

Table IX(b)

Trial No.	F lb/hr	H_1 lb/lb	H_2 lb/lb	S_1 Btu/lb	S_2 Btu/lb	G lb/hr	W lb/hr
89	9.1	0.0140	0.0190	0.247	0.249	1380	7.30
90	9.9		.0200				7.94
91	8.5		.0190				6.82
92	7.7		.0185				5.80
93	6.0	0.0170	0.0211	0.248	0.250	1170	4.95
94	7.9		.0223				6.45
95	6.0		.0212				4.95
96	5.5		.0207				4.55
97	4.7	0.0200	0.0240	0.249	0.251	980	3.91
98	6.0		.0250		.251		4.95
99	3.5		.0228		.250		2.92
100	2.3		.0216		.250		1.93

$17\frac{1}{2}$% LIGNOSOL-VARIATION OF NOZZLE DEPTH AND DRYING GAS VELOCITY

Table IX(c)

Trial No.	λ W.B. Btu/lb	Q_1 Btu/hr	Q_2 Btu/hr	Q_V Btu/hr	Q_F Btu/hr	Q_L Btu/hr	E_D %
89	1027	95,000	71,900	7,500	-224	15,376	7.91
90		93,400	69,800	8,150	242	15,208	8.75
91		94,000	70,500	7,000	208	16,292	7.46
92		93,400	72,500	5,940	178	14,772	6.39
93	1026	79,500	59,000	5,070	-143	15,287	6.40
94		80,500	58,000	6,600	181	15,719	8.22
95		80,000	58,500	5,070	139	16,291	6.35
96		80,000	60,000	4,660	115	15,225	5.85
97	1026	67,100	48,500	4,000	- 95	14,505	5.97
98		68,400	48,500	5,070	120	14,710	7.44
99		68,400	51,500	3,000	71	13,829	4.40
100		66,600	51,500	1,980	47	13,073	2.97

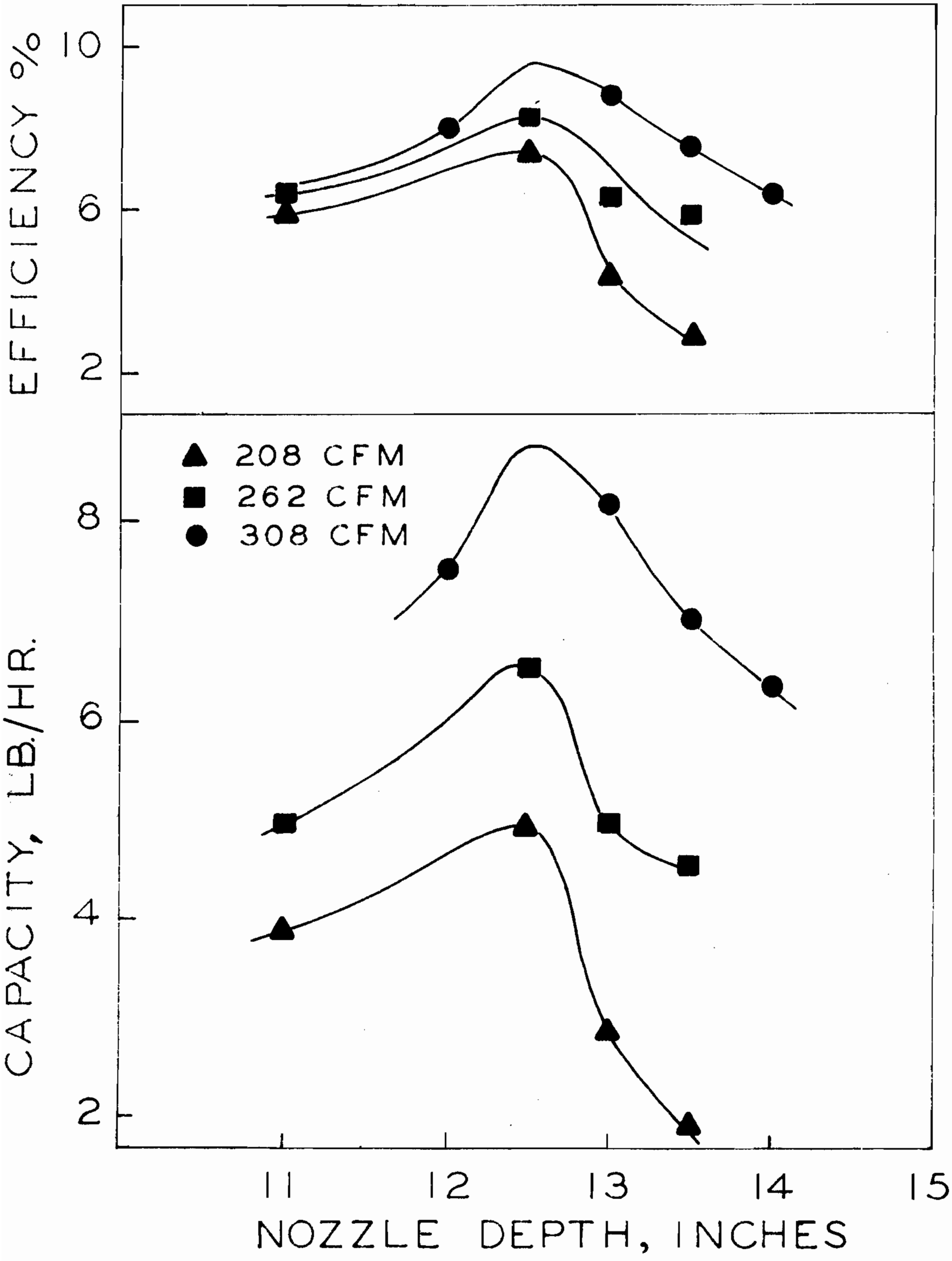

FIGURE 15

17½% LIGNOSOL - VARIATION OF CAPACITY AND EFFICIENCY WITH NOZZLE DEPTH

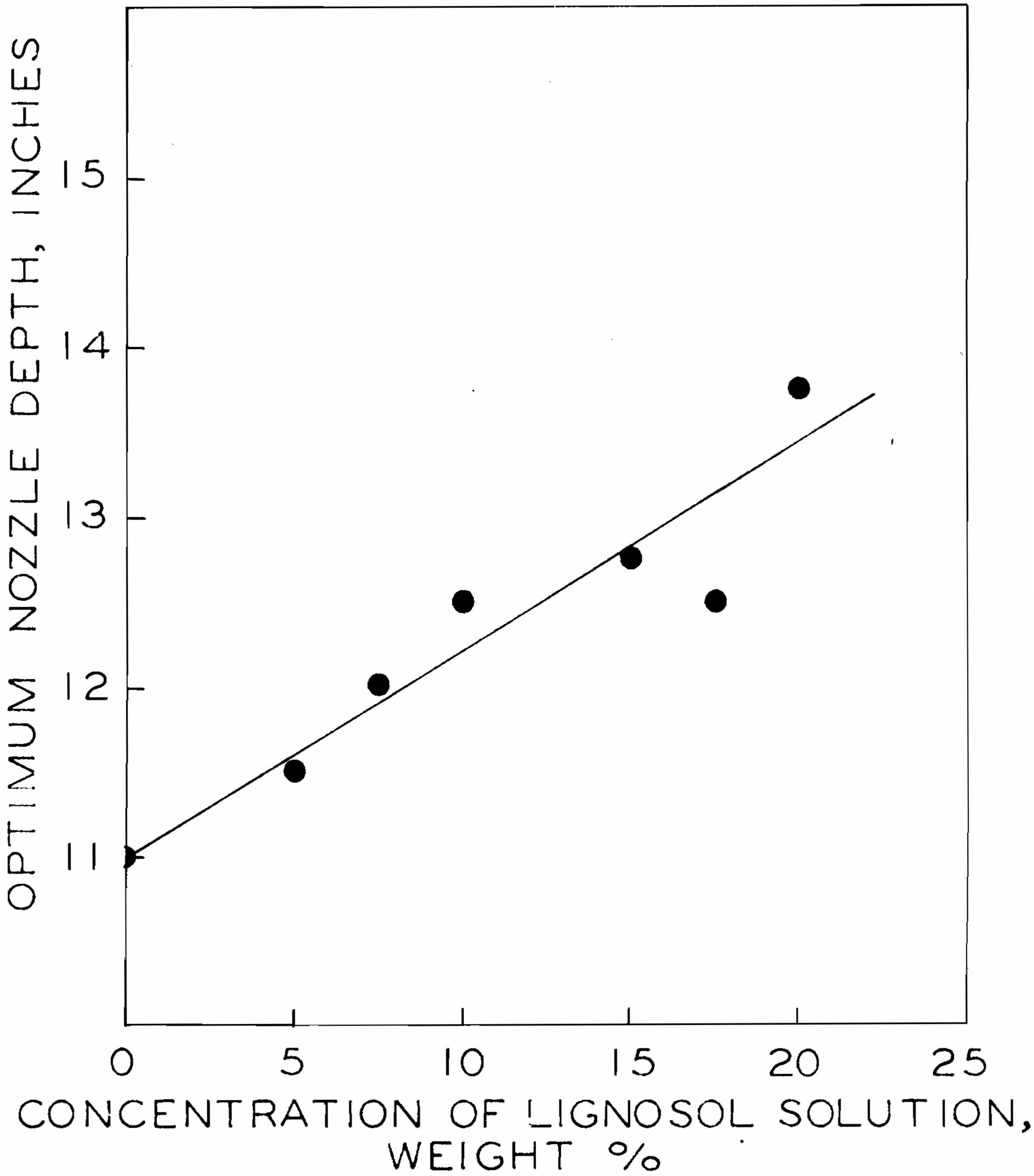

FIGURE 16

VARIATION OF OPTIMUM NOZZLE DEPTH WITH CONCENTRATION

Although the effects of the drying gas rate on capacity and efficiency are inconclusive, a few inferences can be drawn from the experimental results presented. Examination of figures 11 - 15 reveals that each curve of a set has essentially the same shape, regardless of the rate of flow of the drying gases. From this, it would appear that the gas velocity in the drying chamber has little effect on the optimum nozzle position.

From a consideration of the mechanism of spray drying, it is to be expected that a maximum in the capacity curve should be found at a well-defined rate of flow of the drying gas. If the gas moved too slowly, the rate of drying would be too small, causing wet droplets to hit the walls in the lower conical section, where the centrifugal forces are greater. Similarly, at too high a gas rate, the spray would be carried to the wall by centrifugal force before being thoroughly dried. In three of the five cases studied, namely, for pure water and for 7½ and 10 per cent Lignosol concentrations, this maximum occurred within the velocity range covered.

(b) Variation of Drying Gas Temperature

The effect of varying the temperature of the drying gas was investigated with the nozzle depth and gas rate at their optimum values, that is, those corresponding to maximum capacity, for the concentrations being used. The results are shown in tables X - XIV. Figure 17 summarizes the effects on the capacity of the chamber, and figure 18 shows the effects on the efficiency. Good agreement with Knelman's work can be found in the variation of capacity with drying

WATER-VARIATION OF INLET DRYING GAS TEMPERATURE

Table X(a)

Trial No.	l in.	V cfm	T_1 °F	T_F °F	T_2 °F	T_{WB} °F	P_A psig.	P_L psig.
24	11	262	317	80.6	122	107.1	37	34
25			376	89.6	147	115.7	36	34
26			423	91.4	162	122.0	36	37
27			469	95.0	180	123.8	36	37

Table X(b)

Trial No.	F lb/hr	H_1 lb/lb	H_2 lb/lb	S_1 Btu/lb	S_2 Btu/lb	G lb/hr	W lb/hr
24	44.0	0.0150	0.0524	0.247	0.264	1170	44.0
25	47.9	.0190	.0598	.249	.267		47.9
26	51.5	.0250	.0691	.251	.271		51.5
27	54.6	.0350	.0814	.256	.277		54.6

Table X(c)

Trial No.	λ W.B. Btu/lb	Q_1 Btu/hr	Q_2 Btu/hr	Q_V Btu/hr	Q_F Btu/hr	Q_L Btu/hr	E_D %
24	1032	60,600	4,600	41,600	-1,070	13,330	70.0
25	1028	75,600	10,080	49,100	-1,223	15,197	66.1
26	1024	88,500	12,800	52,700	-1,578	21,422	60.7
27	1023	100,900	18,000	56,000	-1,578	25,322	56.5

5% LIGNOSOL-VARIATION OF INLET DRYING GAS TEMPERATURE

Table XI(a)

Trial No.	l in.	V cfm	T_1 °F	T_F °F	T_2 °F	T_{WB} °F	P_A psig.	P_L psig.
43	11.5	308	289	91.4	196	105.8	35	38
44			329	85.1	216	109.4	36	39
45			349	86.0	237	114.8	35	39
46			439	84.2	320	122.0	35	39

Table XI(b)

Trial No.	F lb/hr	H_1 lb/lb	H_2 lb/lb	S_1 Btu/lb	S_2 Btu/lb	G lb/hr	W lb/hr
43	21.4	0.0160	0.0307	0.247	0.254	1380	20.60
44	22.8	.0125	.0286	.246	.253		22.00
45	22.8	.0175	.0337	.248	.255		22.00
46	16.3	.0200	.0315	.249	.254		15.79

Table XI(c)

Trial No.	λ W.B. Btu/lb	Q_1 Btu/hr	Q_2 Btu/hr	Q_V Btu/hr	Q_F Btu/hr	Q_L Btu/hr	E_D %
43	1033	62,750	31,400	21,300	-296	9,754	34.1
44	1032	74,500	37,200	22,750	535	14,015	30.7
45	1028	80,100	43,500	22,600	635	13,365	28.4
46	1024	109,000	70,000	16,180	595	22,205	14.9

7½% LIGNOSOL-VARIATION OF INLET DRYING GAS TEMPERATURE

Table XII(a)

Trial No.	l in.	V cfm	T_1 °F	T_F °F	T_2 °F	T_{WB} °F	P_A psig.	P_L psig.
62	12	262	329	87.8	204	111.2	36	29
63			372	93.2	230	116.6	35	28
64			399	86.9	241	119.3	37	28
65			462	84.2	273	125.6	39	29

Table XII(b)

Trial No.	F lb/hr	H_1 lb/lb	H_2 lb/lb	S_1 Btu/lb	S_2 Btu/lb	G lb/hr	W lb/hr
62	19.8	0.0210	0.0368	0.250	0.257	1170	18.31
63	24.0	.0215	.0410	.250	.259		22.20
64	24.0	.0210	.0400	.250	.258		22.20
65	21.8	.0325	.0500	.255	.263		20.15

Table XII(c)

Trial No.	λ W.B. Btu/lb	Q_1 Btu/hr	Q_2 Btu/hr	Q_V Btu/hr	Q_F Btu/hr	Q_L Btu/hr	E_D %
62	1030	63,750	28,000	18,900	-430	16,420	29.85
63	1027	74,900	34,400	22,800	520	17,180	30.70
64	1026	82,000	37,400	22,800	720	21,080	28.05
65	1022	104,000	46,200	20,600	834	36,366	19.92

10% LIGNOSOL-VARIATION OF INLET DRYING GAS TEMPERATURE

Table XIII(a)

Trial No.	l in.	V cfm	T_1 °F	T_F °F	T_2 °F	T_{WB} °F	P_A psig.	P_L psig.
84	12.5	262	315	80.6	226	109.4	35	30
85			379	84.2	260	116.6		30
86			412	80.6	279	123.8		30
87			446	82.4	306	127.4		31
88			486	89.6	358	131.0		31

Table XIII(b)

Trial No.	F lb/hr	H_1 lb/lb	H_2 lb/lb	S_1 Btu/lb	S_2 Btu/lb	G lb/hr	W lb/hr
84	11.6	0.0163	0.0254	0.247	0.252	1170	10.42
85	15.0	.0210	.0327	.250	.255		13.50
86	17.0	.0308	.0440	.254	.260		15.30
87	16.2	.0362	.0486	.257			14.60
88	12.9	.0370	.0470	.257			11.60

Table XIII(c)

Trial No.	λ W.B. Btu/lb	Q_1 Btu/hr	Q_2 Btu/hr	Q_V Btu/hr	Q_F Btu/hr	Q_L Btu/hr	E_D %
84	1031	59,300	35,000	10,780	-300	13,020	18.22
85	1028	76,900	42,800	13,880	437	19,783	18.12
86	1023	85,500	47,600	15,690	661	21,549	18.50
87	1016	97,500	55,000	14,800	789	26,911	15.30
88	1019	106,800	70,000	11,800	480	24,520	11.10

17½ LIGNOSOL-VARIATION OF INLET DRYING GAS TEMPERATURE

Table XIV(a)

Trial No.	l in.	V cfm	T_1 °F	T_F °F	T_2 °F	T_{WB} °F	P_A psig.	P_L psig.
101	12.5	308	325	79.7	270	109.4	48	41
102			376	80.6	304	117.5		
103			439	85.1	356	121.1		

Table XIV(b)

Trial No.	F lb/hr	H_1 lb/lb	H_2 lb/lb	S_1 Btu/lb	S_2 Btu/lb	G lb/hr	W lb/hr
101	6.0	0.0158	0.0193	0.247	0.249	1380	4.95
102	7.9	.0220	.0270	.250	.252		6.55
103	6.0	.0240	.0244	.250	.251		4.95

Table XIV(c)

Trial No.	λ W.B. Btu/lb	Q_1 Btu/hr	Q_2 Btu/hr	Q_V Btu/hr	Q_F Btu/hr	Q_L Btu/hr	E_D %
101	1031	71,900	54,800	5,100	-147	11,853	7.11
102	1026	86,900	64,600	6,710	242	15,348	7.75
103	1025	105,900	81,400	5,070	496	18,934	4.81

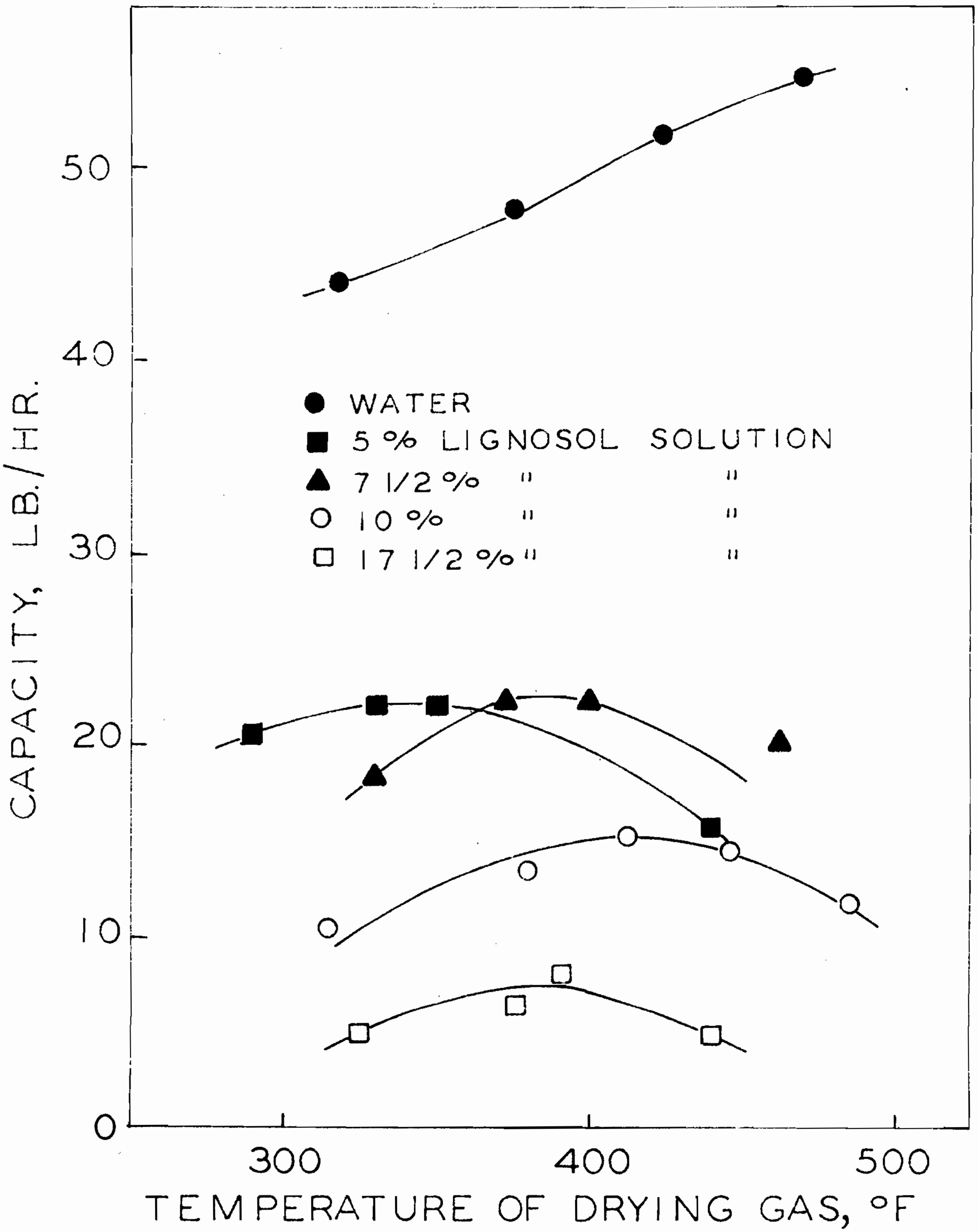

FIGURE 17

VARIATION OF CAPACITY WITH
DRYING GAS TEMPERATURE

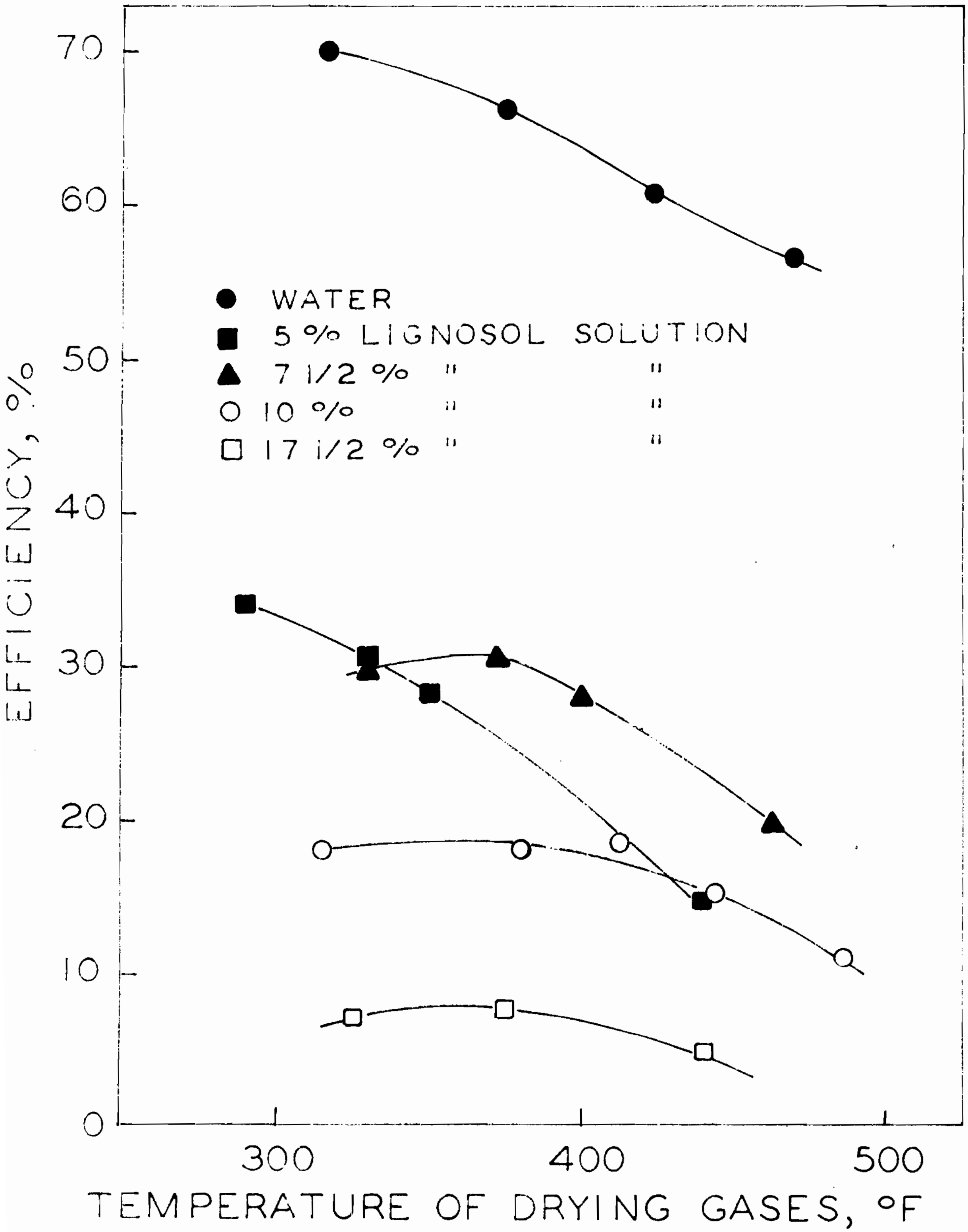

FIGURE 18

VARIATION OF EFFICIENCY WITH DRYING GAS TEMPERATURE gas temperature when pure water is used as the feed liquid. The capacity increases steadily as the temperature level is raised, as would be expected from the simple principles of evaporation which apply when the formation of dried particles is not involved. When Lignosol solutions are dried, however, not only does the capacity of the chamber drop sharply, but, for all concentrations studied, the capacity curve passes through a well-defined maximum at a certain optimum gas temperature. This is an indication that, although simple evaporation still applies during the initial stages, the much slower final stage of drying -- the falling rate period -- now exerts a controlling influence, which predominates more and more as the feed concentration increases. The great reduction in capacity experienced in the presence of more concentrated feeds is rather strikingly demonstrated in figure 17.

The efficiency curves shown in figure 18 again reveal a sharp decrease as the feed is changed from pure water to increasingly more concentrated solutions. It also indicates that use of high gas temperatures is definitely inefficient, owing to higher heat losses and more particularly to poorer utilization of the heat contained in the inlet drying gas, and resulting in progressively higher temperatures of the exit gas.

(c) Effect of Feed Concentration

The experimental investigation of the effects of this important variable has been presented in Tables V - XIV. However, interpretation of the data as plotted in figures 11 - 15 is rather difficult. The relationship becomes much clearer if the maximum capacity of the chamber is compared for each concentration, under otherwise identical operating conditions. Figure 19 shows such a plot at an average inlet gas temperature of 400°F., constant gas rate of 308 c.f.m. and optimum nozzle depth. This graph reveals much more clearly than was apparent from figures 11 - 15 the effect of feed concentration on the drying capacity. The data appear reliable enough to allow interpolation and, to a certain extent, extrapolation.

Of even greater importance, from a practical point of view, is the yield of dried powder per unit time. That greater evaporative capacity for a given set of conditions does not necessarily mean greater production capacity is clearly shown from figure 20, recalculated from trials No. 11, 30, 49, 69 and 90, tables V to IX, and showing the yield of dried powder in pounds per hour at various concentrations. A maximum production of 1.8 pounds per hour was obtained for a feed concentration of 15 per cent, the yield falling off sharply on either side of this optimum point. The values plotted for 20 and for 25 per cent Lignosol solutions were obtained from additional observations which have not been formally presented with the spray drying data.

Finally, the efficiencies of the various runs were plotted against the concentrations on figure 21. The curve, of course, closely follows that of capacity vs. concentration, because the only variable in the efficiency expression is the rate of evaporation of water.

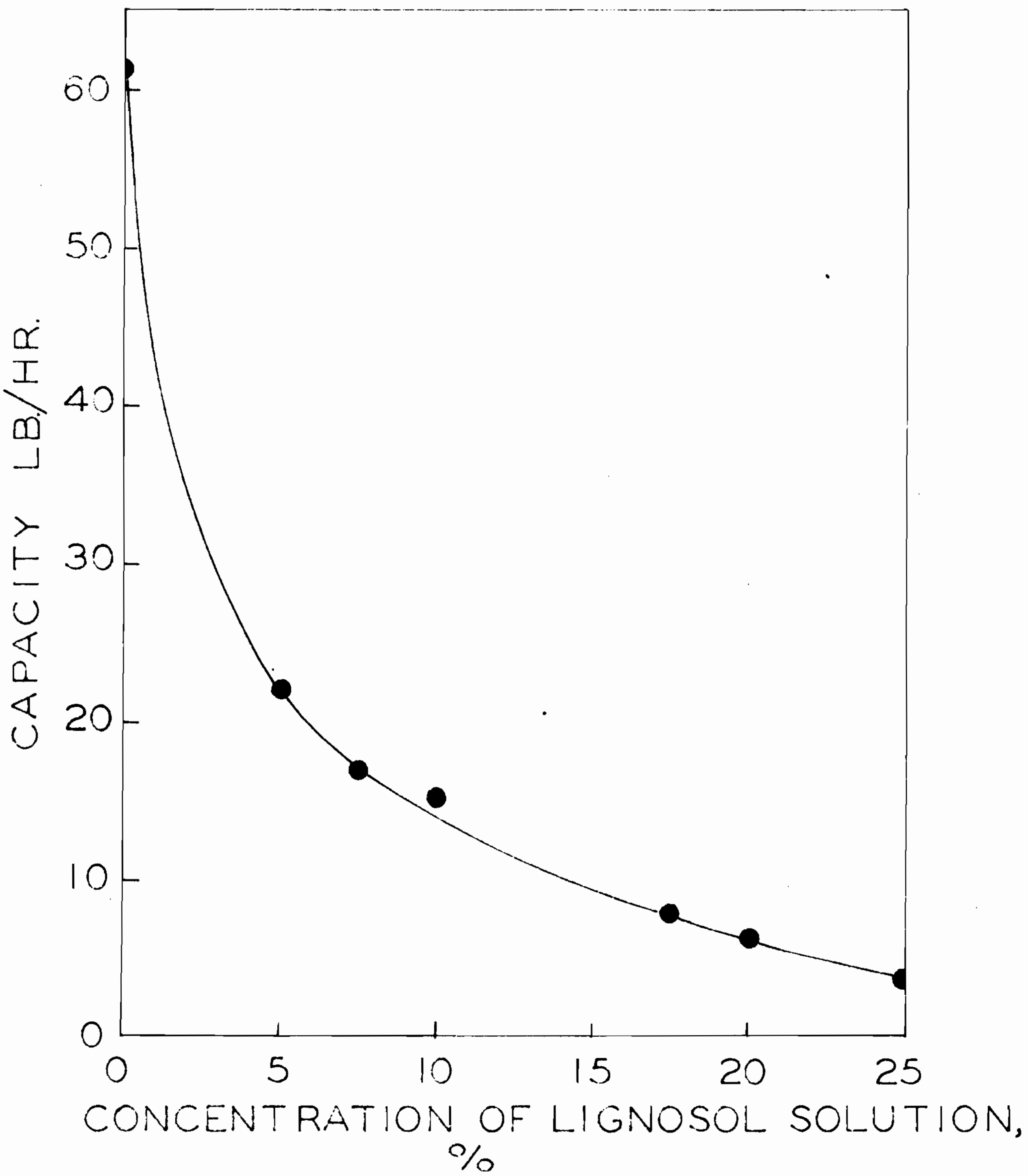

FIGURE 19

VARIATION OF CAPACITY WITH CONCENTRATION

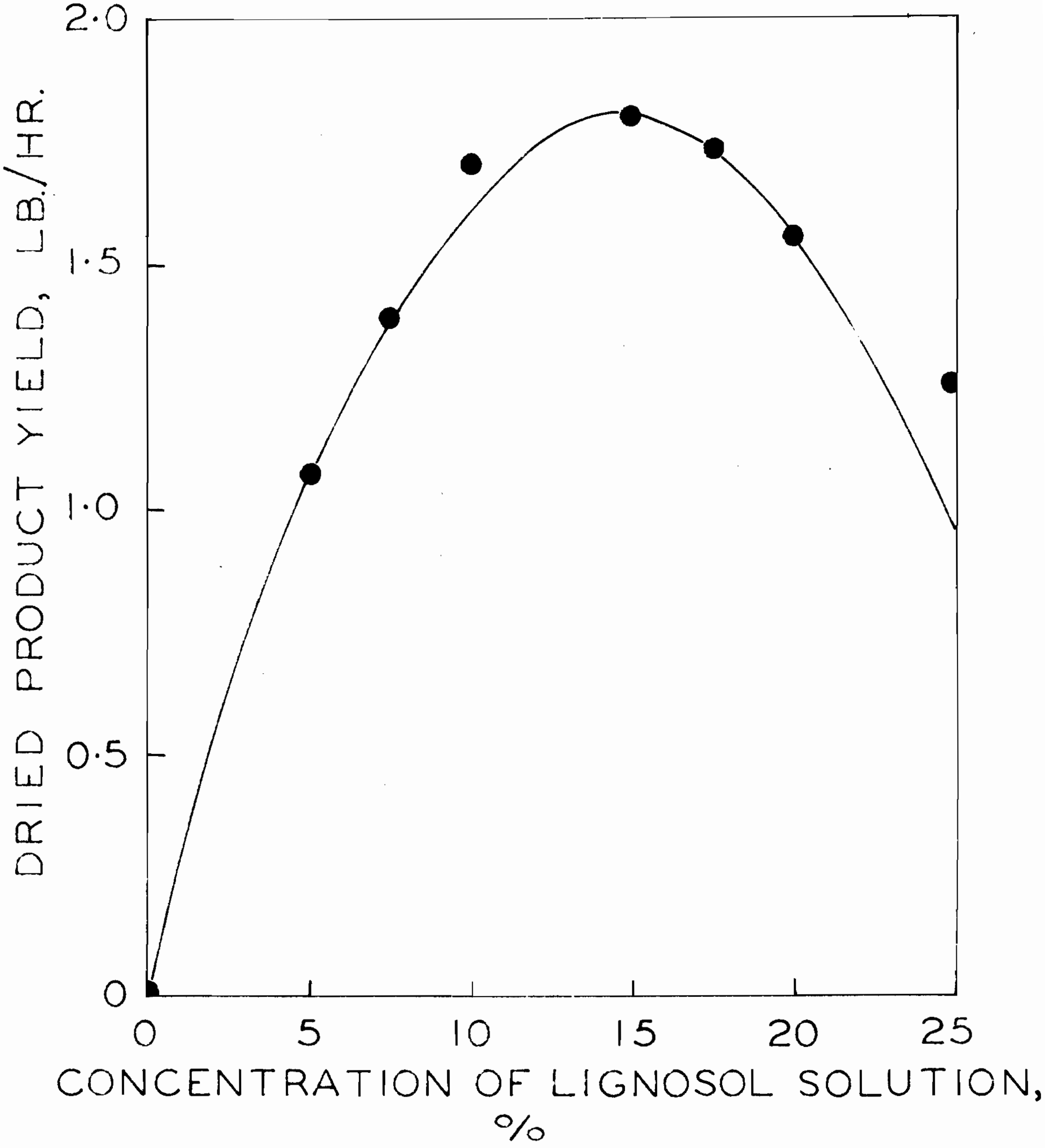

FIGURE 20

VARIATION OF PRODUCTION RATE
WITH CONCENTRATION

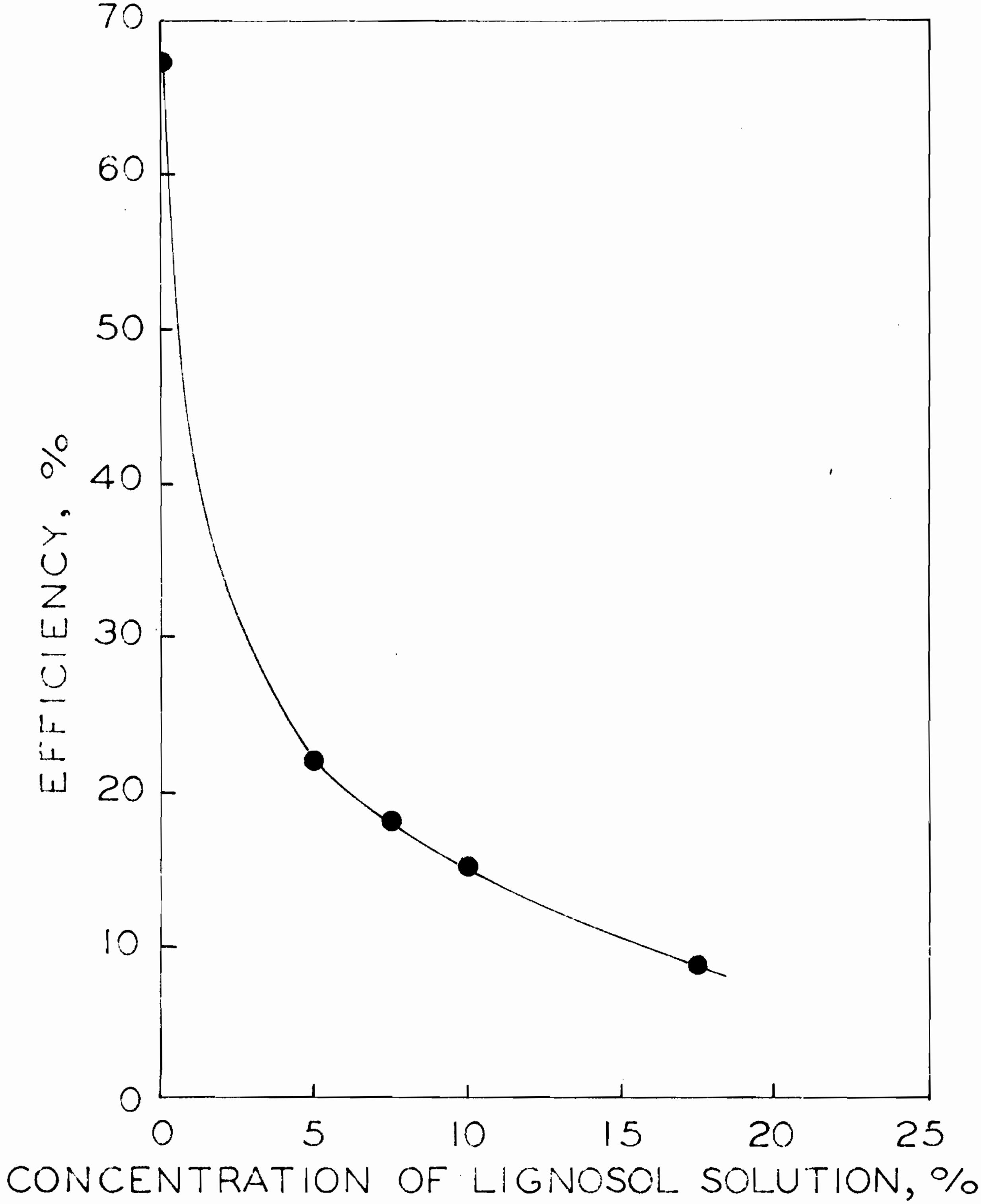

FIGURE 21

VARIATION OF EFFICIENCY WITH CONCENTRATION

IV. ATOMIZATION

Knelman's work has repeatedly emphasized the importance of atomization on the various spray-drying processes. A knowledge of the droplet size obtained with a given set of atomizing conditions is instrumental in understanding the complex mass and heat transfer inter-relationships occurring during such processes. Since no data had been previously published regarding the manner in which the droplet size of the atomized feed and the particle diameter of the dried product would change with feed concentration, a set of runs was made in which the concentration of the Lignosol feed solution was the only variable.

1. EQUIPMENT AND PROCEDURE

The actual spray drying runs were carried out as previously described, the dry product being collected at the Aerotec unit. In order to calculate the required droplet and particle diameters, certain physical properties of the Lignosol were required, namely: the density, viscosity and surface tension of the feed solutions, and the particle size distribution of the dried product. These quantities were obtained by standard physical chemical methods: the density, by the use of a pycnometer; the viscosity, by means of an Ostwald viscometer; and the surface tension, by the capillary rise method. The particle size distribution was determined by the sedimentation technique, using benzene for the suspension medium. Table XV gives the values of these various physical properties over the range of feed concentrations and temperatures covered in this investigation.

Table XV

PHYSICAL PROPERTIES OF LIGNOSOL SOLUTIONS

Lignosol Concentration %	80.6 °F			89.6 °F			98.6 °F		
	ρ gm/cc	μ cp	σ dynes/cm	ρ gm/cc	μ cp	σ dynes/cm	ρ gm/cc	μ cp	σ dynes/cm
0	0.9965	0.855	71.66	0.9951	0.768	70.85	0.9934	0.695	70.05
10	1.0306	1.121	50.90	1.0290	0.991	50.10	1.0269	0.870	49.30
12½	1.0452	1.291	49.75	1.0431	1.128	49.00	1.0407	0.984	48.10
15	1.0569	1.444	48.25	1.0549	1.272	47.25	1.0528	1.101	46.50
20	1.0785	1.921	46.50	1.0762	1.675	45.55	1.0737	1.418	44.50
30	1.1278	4.605	44.10	1.1259	3.875	43.25	1.1231	3.225	42.23
50	1.2017	43.45	-	1.1994	32.35	-	1.1959	28.03	-

2. EXPERIMENTAL RESULTS

The runs in this series were carried out at a constant nozzle depth of 11.5 inches and a drying gas rate and temperature of 308 c.f.m., and 338°F., respectively. The observed data and the physical constants of the solutions used are listed in table XVI(a), while the calculated values necessary for the determination of D_o, as well as the values of D_o, D_p and the thermal efficiency, are to be found in table XVI(b). Figure 22 shows the effects of plotting D_o and D_p against the concentration of the feed.

D_o increases with the feed concentration in a linear manner. This phenomenon, however, is due to the relationship between the physical constants of the solution, and would not necessarily hold true if another type of feed were used. It will be noted from table XVI that the most important factor affecting D_o is the viscosity of the feed. In the discussion of the mechanics of the spraying operation, it was pointed out that the air jet had to overcome the viscous forces holding the liquid body together before it could be atomized. Thus, for a fixed air velocity, as the viscous forces are increased in magnitude, the degree of atomization would be less, and D_o correspondingly larger. If viscosity was the only factor affecting D_o, the curve would rise in a parabolic manner. However, the decreasing values of the ratio of surface tension to density, and of liquid rate to air rate appear to exert a compensating effect, resulting in a very nearly linear relationship.

It is noted that D_p also approximates a linear increase with

VARIATION OF DROPLET AND PARTICLE DIAMETERS WITH CONCENTRATION

Table XVI(a)

C %	F lb/hr	P_A psig	P_L psig	T_A °F	T_F °F	ρ gm/cc	σ dynes/cm	μ poises
5	21.3	35	35	75	90	1.010	56.20	0.00820
7½						1.019	52.75	.00890
10						1.029	50.10	.00991
17½						1.064	46.35	.01450
25						1.100	44.25	.02460

Table XVI(b)

C %	Q_l cfm	Q_a cfm	v_l m/sec	v_a m/sec	v m/sec	E_D %	D_o microns	D_p microns
5	0.00565	2.84	2.84	312	309	27.0	88.15	20.3
7½	.00560		2.82			26.4	95.13	18.3
10	.00555		2.80			25.6	98.28	23.8
17½	.00540		2.72			23.5	108.99	21.4
25	.00518		2.61			21.4	128.50	31.2

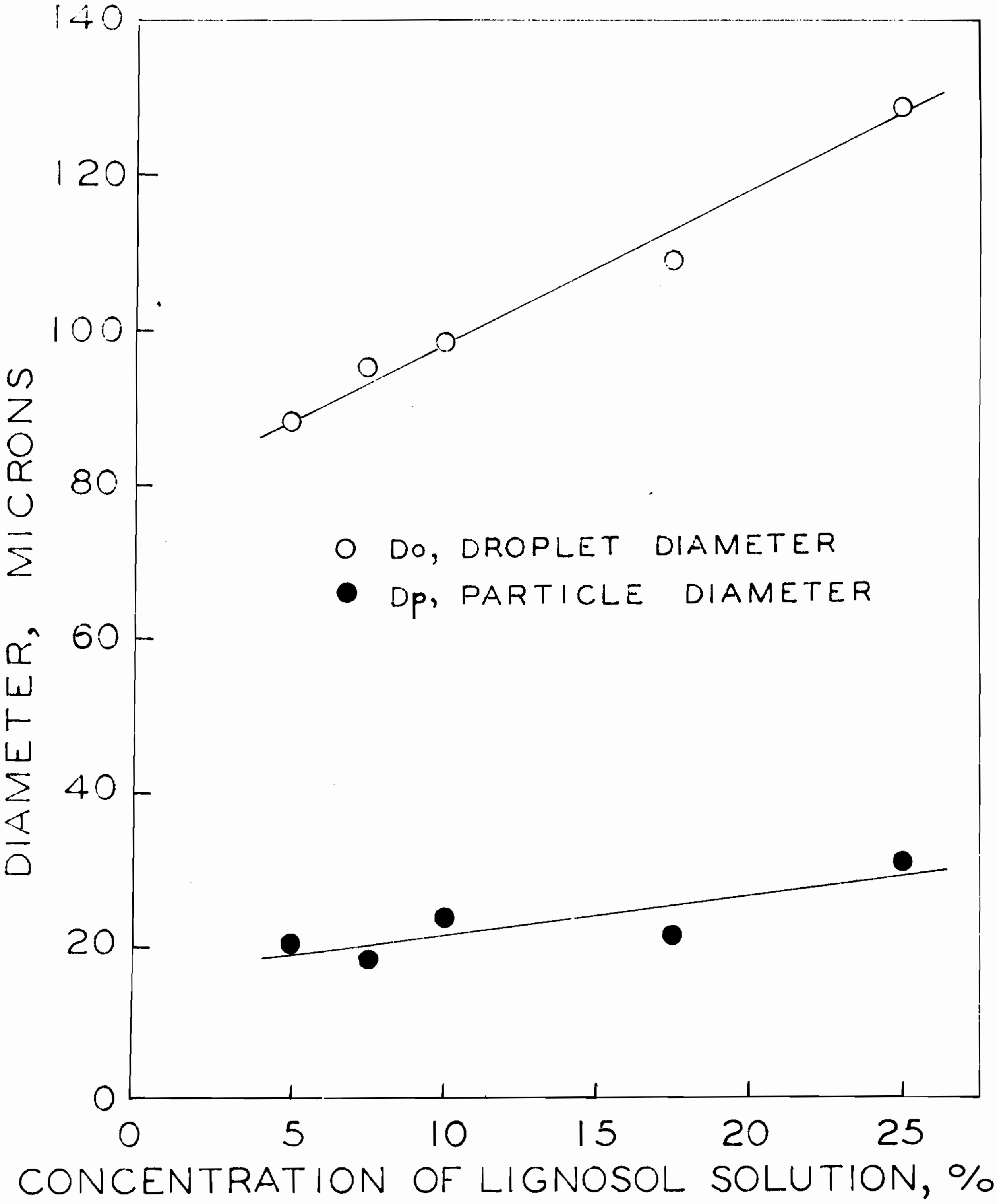

FIGURE 22

VARIATION OF DROPLET AND PARTICLE DIAMETERS
WITH CONCENTRATION

concentration. After a study of the variation of D_o, this result is to be expected, since each droplet contains a higher per centage of solids. This is assuming, of course, that the particle dries in the same manner at each concentration, which was apparently the case over the range studied. The variation of D_p is of interest because of its direct effect on the bulk density, a factor of prime importance in the packaging of commercial spray dried materials.

The equilibrium moisture content of the powder collected at each of the above feed concentrations was obtained using a type C moisture tester supplied by Thos. Ashworth and Company. The results, tabulated below, indicate that the concentration of the feed has little effect on the amount of moisture retained in the product. Similar results have been reported by Chu, Stout and Busche (25) from a study of the spray drying characteristics of Santomerse No. 1.

Feed concentration, weight %:	5	7½	10	17½	25
Moisture content of dried product, weight %:	1.3	1.4	1.35	1.4	1.0

CONCLUSIONS

The experimental results presented in the preceding section amply demonstrate the complexity of the inter-relationships existing between the operating variables in spray drying. However, a somewhat simpler concept of this operation can be obtained if it is considered in the light of its fundamental nature, that is, as a diffusional process. Whether the latter involves transfer of heat, mass or momentum, diffusion will always take place under the influence of a driving force acting to overcome a resistance. For heat transfer, the driving force in spray drying will consist of the temperature difference between the drying medium and the droplet, while the water vapor pressure difference between the same two constituents will represent the driving force for mass transfer. Both are controlled by the temperature and humidity of the inlet gases.

Resistance to diffusion, on the other hand, as encountered in spray drying, should be directly proportional to the thickness of the laminar film surrounding the droplet (a direct function of the relative velocity of the drying medium with respect to the latter) and inversely proportional to the area available for transfer (a function of the degree of atomization). The major role played by this diffusional area -- a minor term in most drying operations -- cannot be overemphasized when the drying of atomized solutions is considered. It has already been pointed out in this connection that a cloud with a total area of 63 square meters can be obtained upon atomization of a litre of water (33).

Since the thickness of the film depends on the relative velocity of the droplet with respect to the drying gas, it thus becomes a function of the rate at which the gases are admitted to the chamber. Unfortunately, this function is not simple, since the velocity of the gas stream is far from being uniform throughout. From gas dynamics, it can be shown that, even in a cylindrical chamber, the tangential velocity of the gas stream is maximum near the walls and decreases rapidly to zero as the vertical axis of the chamber is approached. It is, of course, also zero right at the wall. Thus, a droplet travelling in an orbit of short radius about the vertical axis of the chamber will be exposed to a gas stream of considerably smaller velocity than one travelling closer to the wall. The relationship is further complicated when a conical shape is encountered, since in this case the tangential velocity constantly increases as the radius decreases. It can easily be seen that in chambers of this type, the resistance to diffusion induced by film thickness will be a function of the droplet position, not only in a horizontal plane, but in a vertical plane as well.

Strictly speaking, the drying rate is nothing but the expression correlating the various factors mathematically. It should increase with the driving force, that is, with the temperature of the inlet drying gas; with the degree of atomization, which controls the diffusional area available for transfer; and with the drying gas rate, which decreases the thickness of the film. Actually, of course, it will be affected by the gas stream turbulence as well, since the latter effect is responsible for the removal of the vapor which has diffused through the laminar film, by means of eddy currents.

Finally, consideration must be given to the path available for drying, since the latter will determine, together with the drying rate and the average particle velocity, the actual capacity of the chamber. Increased capacity will result when any of the three factors: drying rate, average particle velocity or particle path, are increased. Maximum capacity will obtain when the length of the particle path is just sufficient to effect the moisture removal at the given rate and under the given conditions of gas flow. It will be exceeded if the path is insufficient.

In the light of the theoretical considerations outlined above, the experimental results will now be examined in some detail.

Measurements of the decrease in temperature of air passing through a two-fluid nozzle revealed that for a pressure of 45 p.s.i.g., the maximum temperature drop obtained was of the order of 8°F. This is in direct contradiction to the work of Fogler and Kleinschmidt (41) who predicted a temperature drop as high as 230°F. for an air pressure of 90 p.s.i.g. These authors calculated the temperature decrease at various pressures by means of the well-known equation for the isentropic flow of an ideal gas:

$$T_2 = T_1\left(\frac{P_2}{P_1}\right)^{\frac{k-1}{k}}$$

where T_1 and P_1, T_2 and P_2 are the temperatures and pressures of the air entering and leaving the nozzle, respectively, and k is the ratio of the specific heats at constant pressure and at constant volume. However, this equation specifies reversibility, which is certainly far from being realized in the expansion of gases through nozzles. A study of the process as plotted on a temperature-entropy diagram shows that, as the degree of irreversibility becomes greater, the temperature drop decreases. Furthermore, these authors ignore the fact that the minimum critical value of the ratio P_2/P_1 is restricted to 0.53 for air, which would limit the theoretical maximum reversible temperature decrease to about 100°F. It is thus evident that the expansion of air through a two-fluid nozzle is highly irreversible, and the large temperature drops predicted by Fogler and Kleinschmidt are not obtained in practice. It was noted that when water alone was sprayed through the nozzle, its decrease in temperature was very small. The cooling effect in this case was undoubtedly due mainly to heat transfer from the liquid to the containing walls, which is supported by the fact that the effect was greater at the higher feed temperatures.

When the liquid feed was atomized in the presence of compressed air, the cooling effect was much more pronounced than in either of the above cases. Since expansion of neither fluid alone can account for this temperature drop, evaporation of the liquid in contact with the air emerging from the nozzle passages appears to offer a more plausible explanation. It should be emphasized that vaporization of only a small percentage of the feed -- which would be rapid due to the highly unstable nature of the initial filaments of liquid formed -- would suffice to provide the cooling effects observed. Further confirmation for these views is supplied by the fact that the spray temperature tends to approach the wet-bulb temperature of the air as a limiting value (see figure 8). These observations are not in complete agreement with the predictions of Edeling (33), who considered that evaporation did not occur to any appreciable extent until the droplet had reached its terminal velocity. However, since these evaporative effects are relatively small, their influence on the droplet trajectory would be negligible, and Edeling's theory of drying would not be invalidated.

For a complete understanding of the marked increase in spray drying chamber capacity with the feed temperature, figure 10, it is necessary to review briefly the path followed by a sprayed droplet, as postulated by Edeling. The droplet, emerging at high velocity from the nozzle outlet, initially follows a trajectory independent of the drying gases. As it approaches its terminal velocity, it is influenced more and more by the drying medium until finally it moves with it in a spiral path down the chamber. According to Edeling, evaporation from the droplet due to the influence of the drying medium will not occur until the droplet has reached its terminal velocity, and until it has reached the wet-bulb temperature of the surrounding medium. From the foregoing discussion, it is readily seen that as the temperature of the feed is increased, the distance which a droplet must travel before reaching the wet-bulb temperature of the drying gases is decreased, yielding a longer drying path and hence a higher capacity. When the feed enters at the wet-bulb temperature, this effect is at a maximum, i.e. the droplet has the longest possible drying path under the existing conditions. At temperatures above the wet-bulb the feed must cool before normal evaporation can take place, which effect is accomplished by a small amount of flashing while the droplet is travelling in its initial trajectory. This effect, however, is small compared to that due to an increase in length of the drying path, and the capacity accordingly rises very slowly with increases in the feed temperature above that of the wet-bulb. A minor factor contributing to the increase of capacity with feed temperature is the fact that D_o decreases as the feed temperature is increased, thus increasing the available diffusional area. The slight decrease of efficiency at the higher feed temperatures is quite easily explained by the fact that, although the heat entering in the drying medium and that used during evaporation are both constant, the heat entering in the feed water is steadily increasing, thus decreasing the thermal efficiency.

One of the most striking results obtained was the critical nature of the nozzle position in the chamber, as shown in figures 11 - 15. That an optimum nozzle depth should exist can readily be demonstrated by a consideration of the relative paths travelled by the droplets and by the drying medium. If the nozzle is located near the top of the chamber in the cylindrical section, the terminal velocity of the spray is reached at a point close to the axis of the chamber, where the velocity of the gas stream is relatively small, as explained above. The droplets will thus assume an orbit of short radius, and the relative velocity of the gas stream will be correspondingly low. Admittedly the latter will increase as the gas spirals down into the conical section, but owing to the low initial drying rate, the droplet will still retain an appreciable bulk and will easily escape to the wall by centrifugal action. Conversely, if the nozzle is located well down in the conical part of the chamber, where the gas velocity is much higher, the centrifugal forces will again predominate and carry the droplets to the wall. Thus, at some point in the chamber between these two extremes, the droplets achieve a maximum length of drying path and hence a maximum capacity. By a very similar mechanism, as the concentration of the feed is raised, both the diameter of the droplet and its mass increase, as shown in the studies of atomization, figure 22. This, in turn, increases the initial momentum and the necessary drying time, and necessitates a progressively lower position of the nozzle in the chamber to allow the droplets to reach their terminal velocities in zones of increasing tangential velocities. The fact that the drying path of the droplet is being decreased, while the necessary drying time is increased, are probably the essential factors responsible for the decrease of capacity with concentration, summarized in figure 19. Here again, the complexity of the factors involved manifests itself: the change in physical properties of the feed causes a larger value of D_o, thus producing a smaller diffusional area with a consequent poorer heat transfer; although the amount of water to be evaporated per unit mass of feed is less, this advantage is more than offset by the slower drying rates in the presence of the higher solid content of the droplet.

Of particular significance is the fact that, at any feed concentration, the optimum nozzle position, as shown in figures 11 - 15, appears to be independent of the drying gas rate, at least, within the range studied. The reason for this interesting generalization is not at once apparent, although it may be due to the fact that, near the vertical axis of the chamber, there exists a zone of more or less quiescent, low velocity gas, which is little affected by the rate of circulation of the bulk of the drying gas, which might be visualized as revolving in an annulus close to the walls. Since the effect of the nozzle position is essentially to make the terminal velocity of the spray fall just outside of this central zone, little dependence on inlet gas rate might be expected.

The variation of the yield of dried powder at optimum capacity with concentration, figure 20, is due simply to the two opposite effects of increased solid content in the feed and decreased capacity of the chamber. The former factor initially predominates as the concentration increases, but the rapid fall in capacity causes the rate of production of dried Lignosol to pass through a maximum and then to decrease.

For the evaporation of pure water, higher drying temperatures should yield faster drying rates and increased capacities, as shown in figure 17. However for a solution, the falling rate period of drying with its attendant phenomena -- bursting, shrinking, surface-hardening, etc., -- becomes the controlling factor, and a maximum drying temperature prevails beyond which no increase in capacity will be obtained for a corresponding increase in temperature. As a matter of fact, the opposite effect was noted in the present investigation, a continued increase in the temperature of the drying medium caused the capacity to decrease slightly after passing through a maximum value. It is probable that in this case intensification of such effects as surface-hardening at the more elevated temperatures caused a decrease in the rate of diffusion of water through the partially dried particle, and thus decreased the drying rate. Even in the absence of such effects, increased heat losses should limit the capacity at higher gas temperatures. The efficiency, of course, decreased steadily at the higher temperatures because increased amounts of heat were entering the chamber without a corresponding rise in evaporation.

The results of this investigation are in excellent agreement with those obtained by Knelman, except as otherwise noted in the body of the report. A comparison of the physical properties of the Lignosol solution used by Knelman, more particularly the density, with the data tabulated on page 110 indicates that the feed concentration used in his investigations was not 10 per cent as reported, but closer to 13 per cent. This error, which should have no effect on the trends reported, came about because the Lignosol used by Knelman was received from the manufacturers in solution form of approximate concentration, whereas that used in the present work was received as a dry powder and the solutions made up to the proper concentration, as required.

Although the data collected do not permit the development of mathematical relationships between the various spray drying variables, it is felt that the qualitative results presented provide the necessary foundation for more fundamental research on the problem. Further work on the subject is definitely required, and should include such topics as investigation of Edeling's predictions regarding the drying time of a droplet or particle, refined methods of calculating the heat transfer coefficient and its relation to the coefficient of mass transfer, and methods of assessing the degree of turbulence existing at any point in the drying chamber.

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