

A MANUAL
FOR THE THERMAL AND HYDRAULIC DESIGN OF DIRECT CONTACT SPRAY
COLUMNS FOR USE IN EXTRACTING HEAT FROM GEOTHERMAL BRINES

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ABSTRACT

This report outlines the current methods being used in the thermal and hydraulic design of spray column type, direct contact heat exchangers. It provides appropriate referenced equations for both preliminary design and detailed performance. The design methods are primarily empirical and are applicable for use in the design of such units for geothermal application and for application with solar ponds. Methods for design, for both preheater and boiler sections of the primary heat exchangers, for direct contact binary powers plants are included.

INTRODUCTION

The spray column has been widely studied in the chemical industry for many years due to its inherent simplicity as a counter-current device for heat or mass transfer. Developments were enhanced in the 1960's due to increased interest in desalination systems.⁽¹⁾ More recently, in the 1970's, Jacobs and Boehm⁽²⁾ suggested their use for extracting heat from moderate temperature geothermal brines. They and a number of other investigators have carried out a wide range of studies under U.S. Department of Energy funding for nearly ten years. This work culminated in the construction of the 500 KW_e Geothermal Direct Contact Binary Cycle Power Plant at East Mesa, California, by Barber-Nichols Engineering under U.S.D.O.E. funding.⁽³⁾ The 500 KW_e direct contactor was designed by the present author as a combined working fluid preheater-boiler. The working fluid was isobutane with the continuous fluid being the immiscible geothermal brine.

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Based upon the relative success of this application, a number of other applications have been spawned. Most closely related is the use of a modified spray type direct contactor for the extraction of heat from a salt stratified solar pond. The low temperature design conditions for a solar pond dictate the use of pentane as the working fluid if it is desired to utilize the working fluid vapor to generate electricity.

Although both geothermal and solar pond applications have the same ultimate purpose, to generate electricity from a moderate to low temperature source and to obtain the energy exchange at small approach temperature differences, many source related characteristics cause significant differences in their design. For the geothermal application, it has generally been conceded that the most economical design is to utilize as much heat as possible from each unit mass of geothermal brine. This leads to near equal mass flow rate of the working fluid and brine. For the case of solar ponds, with much lower peak temperatures and concerns about returning too cold a fluid back to the pond, the mass flow rate of brine far exceeds that of the working fluid.

For a combined boiler-preheater, it is clear that for high-pressure, high-temperature vapor generation (such as for geothermal applications) the heat duty of the preheater can greatly exceed that of the boiler. For solar pond applications, where the vapor is generated at temperatures as low as 67°C , the boiler duty can be two to three times that of the preheater. Thus, design philosophy can be considerably different. Nevertheless, in this design manual, an attempt is made to provide information for general purpose design

of spray column type direct contactors. As nearly as possible, the information provided herein is current and provides the best available information.

SPRAY COLUMN DESCRIPTION

A spray column is one of the oldest known devices for contacting two immiscible fluids in order to transport either heat or some chemical substance from one fluid to the other. It is basically an empty vertical column with injection devices for each fluid and outlets for each fluid. In most common applications, each fluid is in a liquid phase; however, for use with binary power cycles, a single column can include a liquid-liquid preheating zone and a boiling or evaporation zone.

When only liquid-liquid heat exchange is desired, the column must have a disengagement zone at both the top and bottom of the column (see Figure 1). A properly designed column needs only to control the flow rates of the two liquid streams to insure a pseudo-steady operation.

A column in which both preheating and boiling takes place requires that there be a disengagement section at the bottom of the column, a level control device to insure that the column is not completely flooded with liquids, and a vapor reservoir at the top of the column for the generated vapors. The vapor reservoir must be sufficiently large to insure that a liquid phase does not exit as a mist with the vapor mixture. Thus, mist eliminators may also be required (see Figure 2).

DESIGN OF DISPERSED PHASE FLUID INJECTORS

In order to design a direct contact heat exchanger of the spray column type, it is necessary to design a distributor which can produce regular uniform-sized drops of one of the two fluids. Normally this is the lighter fluid. Thus, for geothermal or solar applications, this would be the hydrocarbon working fluid. This is achieved by designing a distributor which uses a perforated plate of a material not wetted by the fluid to be dispersed to form the drops. Typically, punched holes which leave a slight nozzle at the surface exposed to the continuous phase are used. For geothermal applications, a mild steel plate pickled in sulfuric acid provides clean jets and well-formed drops.

Actual design of the holes is not critical as long as the flow rate through them is equal to the jetting velocity, but less than the critical jetting velocity. Exceeding or equaling the jetting velocity is important due to the fact that lower velocities can lead to situations where not all of the nozzles are flowing and due to the fact that drops formed when V_j is exceeded are very regular in size. Regular size drops are important in order to predict column performance. Steiner and Hartland⁽⁴⁾ recommend that the jetting velocity be calculated from

$$V_j = \sqrt{\frac{2\sigma}{\rho_d d_n}} \left(1.07 - 0.75 \sqrt{\frac{\Delta\rho d_n^2 g}{4\sigma}} \right) \quad (1)$$

where σ is the interfacial tension*, d_n is the nozzle diameter.

It is also necessary not to exceed the critical jetting velocity. Above this velocity the length of jet decreases dramatically followed quickly by atomization of the dispersed phase. This requires a large pressure drop across the nozzle and is generally undesirable for spray columns. The critical velocity and corresponding critical drop diameter can be calculated according to Skelland and Johnson⁽⁵⁾ as follows

$$d_{jc} = \frac{d_n}{0.485K^2 + 1} \quad \text{for } K < 0.785 \quad (2)$$

or

$$d_{jc} = \frac{d_n}{1.51K + 0.12} \quad \text{for } K \geq 0.785 \quad (3)$$

where

$$K = \frac{d_n}{\sqrt{\sigma / \Delta \rho g}} \quad (4)$$

$$V_{jc} = 2.69 \frac{d_{jc}^2}{d_n^2} \sqrt{\frac{\sigma}{d_{jc}(0.514\rho_d + 0.472\rho_c)}} \quad (5)$$

*Interfacial tension is not surface tension. It can be predicted by Antonoff's rule which states that for two saturated liquid layers in equilibrium, the interfacial tension is equal to the difference between the two individual surface tensions of the two mutually saturated phases under a common vapor or gas, $\sigma = \sigma_{ds} - \sigma_{cs}$. If the above values are not known saturated-phase surface tensions can be used. Accuracy within 15% is claimed for organic-water and organic-organic systems for the latter estimate. Reference 1 gives values for organic fluids and water and brine under air. These properties should be used.

At this critical velocity, Treybal⁽⁶⁾ recommends the following equations for the critical drop size

$$d_{DC} = \frac{2.07 d_n}{0.485 E_o + 1} \quad \text{for } E_o'' < 0.615 \quad (6)$$

and

$$d_{DC} = \frac{2.07 d_n}{1.51 E_o^{1/2} + 0.12} \quad \text{for } E_o'' \geq 0.615 \quad (7)$$

where E_o'' is the Eötvös number, defined as

$$E_o'' = \frac{\Delta \rho g d_n^2}{\sigma} \quad (8)$$

For conditions between the jetting velocity and the critical jetting velocity, the following correlation is recommended⁽⁴⁾ to determine the drop size.

$$d_D = d_{jc} \left(2.06 \frac{V_{jc}}{V_n} - 1.47 \ln \frac{V_{jc}}{V_n} \right) \quad (9)$$

Reference 4 recommends maintaining a minimum Weber number (defined with nozzle velocity and the density of the dispersed phase,

$$\text{i.e., } W_e = \frac{V_n^2 d_n \rho_d}{\sigma}) \text{ greater than two to prevent seeping along the}$$

surface and secure drop formation on all openings. Experience in the laboratory indicates that nozzle or perforation spacing should not be closer than $1.5 d_D$ to insure that jet or drop coalescence does not occur.

BEHAVIOR OF DROPS

Drops formed from jets or nozzles may behave differently according to their density, interfacial tension, volume, and whether heat or mass transfer takes place between them and the surrounding continuous phase. For a drop rising due to gravity in an immiscible liquid, there are five dimensionless groups that govern the motion of the drop:

$$\text{Reynolds number} \quad Re = \frac{\rho_c d_D V_D}{\mu_c} \quad (10)$$

$$\text{Eötvös number} \quad Eo = \frac{\Delta \rho g d_D^2}{\sigma} \quad (11)$$

$$M - \text{group} \quad M = \frac{g \mu_c^4 \Delta \rho}{\rho_c^2 \sigma^3} \quad (12)$$

$$\text{Viscosity ratio} \quad K_1 = \mu_d / \mu_c \quad (13)$$

$$\text{and Density ratio} \quad \gamma = \rho_d / \rho_c \quad (14)$$

For any particular liquid-liquid combination M , K , γ are constant in an isothermal system. Thus, Grace⁽⁷⁾ correlated drop behavior by plotting Re versus Eo for constant values of M for a large number of liquid pairs. K_1 and γ play a small role in the results. Figure 3 categorizes drop into three regimes: the Spherical regime, the Ellipsoidal regime and the Spherical Cap regime. An approximate curve is shown which separates the former two regimes. (Experiments conducted at the University of Utah at high values of Re for Eo of near one indicate that the spherical regime exists longer than shown.)

The spherical regime contains that region where drops are spherical, or nearly so. For spherical drops, little or no internal circulation takes place.

Somewhat larger drops obtain, on a mean time basis, a shape like that of an oblate ellipsoid of revolution. The instantaneous shape may depart radically and undergo wobbling which, of course, would cause significant internal circulation.

When $E_o \geq 40$ for all $M \leq 10^2$ droplets have a leading surface which looks spherical, but the rear may be flat or concave. These drops may move randomly and their behavior is hard to correlate. Thus, they should be avoided in the design of a spray column.

For spherical drops, the terminal velocity in a quiescent fluid can be calculated by a simple balance of the gravity force by the drag yielding

$$f V_T^2 = \frac{4}{3} d_D \left| \left(\frac{\rho_d - \rho_c}{\rho_c} \right) \right| g \quad (15)$$

If it is assumed that the drops behave like rigid smooth spheres, then f varies as follows

$$\begin{array}{ll} \text{Re}_c < 0.1 & f = 24/\text{Re}_c \\ 2 < \text{Re}_c < 500 & f = 18.5/\text{Re}_c \\ 500 < \text{Re}_c < 2 \times 10^5 & f = 0.44 \\ 2 \times 10^5 < \text{Re}_c & f = 0.2 \end{array} \quad (16)$$

More recently Rivkind and Ryskin⁽⁸⁾ have proposed for the drag coefficient:

$$f = \frac{1}{K+1} \left[K \left(\frac{24}{\text{Re}_c} + \frac{4}{\text{Re}_c^{2/3}} \right) + \frac{14.9}{\text{Re}_c^{0.78}} \right] \quad (17)$$

which accounts for the relative motion of the interface due to the differences in fluid viscosities. It should be noted, however, that the presence of contaminants such as found in geothermal brines or salt pond brines tend to make the interface more immobile. However, data is missing on the influence of surfactants and impurities.

For ellipsoidal drops Grace⁽⁷⁾ recommends that the terminal drop velocity be calculated from

$$V_T = \frac{\mu_c}{\rho_c d_D} M^{-0.149} (J - 0.857) \quad (18)$$

where

$$J = 0.94H^{0.757} \text{ for } 2 < H < 59.3 \quad (19)$$

$$J = 3.42H^{0.441} \text{ for } H > 59.3 \quad (20)$$

with

$$H = \frac{4}{3} E_o M^{-0.149} \left(\frac{\mu_c}{0.0009} \right)^{-0.14} \quad (21)$$

In the above equations μ_c is in kg/msec.

VELOCITY OF DROPS IN SWARMS

Drops in a spray column, depending upon the holdup, may move in dense swarms. As the drops get closer together they interact changing not only their own velocities but also that of the continuous phase. Steiner and Hartland⁽⁴⁾ recommend

$$\frac{G_c}{A} = \frac{\left[\frac{k \Delta \rho d_D g}{\rho_c} \right]^{1/2} (1 - \epsilon)^{1+n/2}}{(1 + \epsilon^{1/2})^{n/2} \left(1 + \frac{R(1-\epsilon)}{\epsilon} \right)} \quad (22)$$

to predict the superficial velocity of the continuous phase. This compares with

$$\frac{G_c}{A} = \frac{V_T \epsilon (1-\epsilon)^{m+1}}{(1-\epsilon)R + \epsilon} \quad (23)$$

which has been used by Letan, et al.^(9,10) and by the present author⁽²⁾. In Equation 22, recommended values for k and n are 2.725 and 1.834 respectively. The value of m in Equation 23 is a function of the drop Reynolds number based on terminal velocities, as follows:

$$\begin{aligned} \text{Re}_c < 0.2 & \quad m = 3.65 \\ 0.2 < \text{Re}_c < 1.0 & \quad m = 4.35 \text{Re}_c^{-0.03} - 1 \\ 1 < \text{Re}_c < 500 & \quad m = 4.45 \text{Re}_c^{-0.1} - 1 \\ \text{Re}_c \geq 500 & \quad m = 1.39 \end{aligned} \quad (24)$$

In the above equations, ϵ is the holdup.

In low holdup situations, the relative velocity between the drops and the continuous phase is equal to the terminal rise velocity of a single drop.

$$V_r = V_{D_T} = \frac{G_c}{A(1-\epsilon)} + \frac{G_d}{A\epsilon} \quad (25)$$

However, at higher holdup the close proximity of adjacent drops influences the terminal velocity of a typical drop within the swarm, thus, $V_r \neq V_{D_T}$. Kumar⁽¹¹⁾ recommends for this situation that

$$V_r = \left[\frac{2.725 \Delta \rho d_D g}{\rho_c} \left(\frac{1-\epsilon}{1+\epsilon^{1/2}} \right)^{1.834} \right]^{1/2} \quad (26)$$

It was this expression together with Equation 25 from which Equation 22 was developed. Comparison of Equation 26 with experimental data from other investigators indicate mixed results especially at low holdup. Nevertheless, it is recommended by Steiner and Hartland⁽⁴⁾. Deviation from $V_r = V_T$, however, is small unless dense packing is achieved. Thus, Equation 23 is preferred unless high holdups are encountered.

HOLDUP OF THE DISPERSED PHASE

The holdup is the fraction of the total volume occupied by the dispersed phase. In an isothermal spray column, it is constant along the column length. However, when heat transfer occurs, the holdup can vary along the column length. Whereas for many applications the variation may be small, in a geothermal application using isobutane as the working fluid, changes in holdup must be taken into account in the preheater much less the boiler (if a boiler-preheater combination is used). This is due to the fact that density changes of 25% can readily occur in the preheater.

If the continuous phase flow rate is held constant and the dispersed phase flow rate is gradually increased, a condition known as flooding can eventually occur. (Similarly, flooding will eventually occur if the dispersed phase flow rate is held constant and the continuous phase flow is increased.) At this point, it is impossible to continue passing more of the dispersed fluid through the column and a fraction would then be washed out of the column with the continuous phase or the drops might all coalesce, changing either the drop size or which fluid is dispersed. As this cannot be readily predicted,

flooding is to be avoided. We, thus, must be able to predict the flooding point.

There is not much data on flooding in the literature⁽⁴⁾. Based on the relationship of Richardson and Zaki⁽¹²⁾, Letan⁽¹⁰⁾ recommends that

$$(m+1) (1-R) \epsilon_f^2 + (m+2) R \epsilon_f - R = 0 \quad (27)$$

be used to calculate the holdup at flooding where m is given in Equation 24 and $R = G_d/G_c$, the ratio of volumetric flow rates. For stable flow operation

$$\epsilon < 0.9 \epsilon_f \quad (28)$$

is advised.

It should be noted, that it is possible to operate at a holdup greater than the flooding point. This type of operation occurs in what is called the dense packing regime. Operating below the so-called flooding point is called the dispersed packing regime. Theoretically it is possible to operate a spray column in each regime for a given pair of flow rates. Although considerable work has been directed toward dense packing, in practice it is difficult to achieve. It does have advantages over the common operation with loose packing. The interfacial area can be three to five times higher and back mixing is considerably reduced. The dense dispersion can be controlled so that its interface is 30-50 mm above the dispersed phase distributor. However, if the jets from the distributor plate enter into the dense packing, it can lead to coalescence of the drops. Once coalescence starts, it continues through the column height and operation cannot be

maintained. This coalescence is the main danger in operating in the dense packing regime.

Not much is known about either the hydrodynamics or the heat transfer in the dense packing regime. Thus, although it is possible to operate in this manner, prudent designers have stuck to disperse packing operation. Further, research to counteract coalescence by the adding of surfactants to the drops could result in practical dense phase operation of a spray column. The influence of the surfactants on heat and mass transfer would also have to be studied.

AXIAL MIXING

Spray columns are designed with the intent to facilitate countercurrent contact between two immiscible fluids. The degree to which this is achieved depends upon the design of the injectors for the two fluids. The design for the dispersed phase is reasonably straightforward and is essentially a perforated plate covering a manifold. The limiting factors on nozzle jet velocity have already been discussed. The injection of the continuous phase is more difficult and flaws in its design are probably the leading cause of axial circulation in a liquid-liquid spray column. However, it should be noted that little research on internal back mixing or axial mixing has been done.

In the analyses of spray columns, models used are normally one-dimensional and transient or steady state. Back mixing can be introduced by introducing single dispersion coefficients, E_c , and E_d which may be correlated by comparing column operating results against different operating parameters. Principally, it is assumed that an

additional flux exists in the opposite direction to the main flow of each of the phases. The magnitude of the back flow is assumed to be proportional to the negative gradient of the parameter in question (temperature, concentration, velocity, etc.). For example, for mass transfer, the conservation equations take the form⁽⁴⁾

$$\frac{\partial x}{\partial t} = V_c \frac{\partial x}{\partial z} + E_c \frac{\partial^2 x}{\partial z^2} - \frac{k_c a}{1-\epsilon} (x-x^*) \quad (29)$$

for the continuous phase, and

$$\frac{\partial y}{\partial t} = -V_d \frac{\partial y}{\partial z} + E_d \frac{\partial^2 y}{\partial z^2} + \frac{k_c a}{\epsilon} (x-x^*) \quad (30)$$

for the dispersed phase. The values for E_c and E_d will be dependent on column geometry, injector design, etc., as well as the fluids being used. These points are discussed by Reference 4. They carried out experiments in a spray column without any dispersed phase present. They noted strong circulation in the continuous phase only. Other back mixing can be caused by the fact that the drops carry wakes that are periodically shed as they rise. As pointed out by Letan⁽⁹⁾, in a sufficiently tall column, this effect can even out along the length of the column. In her heat transfer work this resulted in long columns operating with a nearly constant value of the volumetric heat transfer along the column length for columns with length to diameter ratios, L/D_c , greater than eight.

In small diameter columns, bulk circulation can occur due to the fact that drops concentrate in the middle of the column. In larger diameter columns, this can lead to more difficulties as it has been proposed that the axial dispersion is proportional to the diameter raised to a power ($E \sim D^n$). However, it is believed that this is

again caused by the difficulty in designing an appropriate continuous phase inlet. Such was the problem, it is believed, with the 500 KW_e spray column direct contactor at East Mesa. However, changes in other operating conditions were also made at the same time as the injector was changed. Thus, no one knows for sure whether the injector modification alone led to improved operation. However, testing with continuous phase nozzle design prototypes at the University of Utah in its six inch diameter unit, seemed to indicate significant reduction in back mixing. Thus, it is believed that appropriate design can significantly reduce back mixing.

No general rule is available to design the continuous phase injector. The following rules of thumb are proposed:

- 1) The continuous phase injector should not release a strong jet of fluid. A local strong jet produces recirculation regions about it. If the fluid is injected downward, strong axial recirculation cells may develop. Further, a strong jet can lead to local flooding and potential breakup of the drops.
- 2) Lateral release of the continuous phase is desirable; however, again, strong flows can lead to drop breakup. While this may be tolerated in a boiler-preheater combination where the nozzle is in the boiler region it cannot be tolerated in the preheater where countercurrent flow is strongly desired.
- 3) Multiple spaced inlets can insure low speeds and small velocities, thereby, minimizing any jet lengths and thus recirculation zones. No other guides are available, although examples of inlets are provided in the literature.

HEAT TRANSFER CORRELATIONS

Liquid-Liquid Heat Transfer

Direct contact heat transfer between the drops of the dispersed phase and the continuous phase is complex. It depends not only on the thermal properties of each phase, but also on the dynamics of the drops themselves. As was noted earlier, the drops can be spherical, ellipsoidal or cap shaped depending upon their Reynolds number, their Eötvös number and their M-group. To date, most experiments for spray columns were carried out with drops whose diameters were less than 7.5 mm with most heat transfer studies being carried out for drops with $2.0 \text{ mm} \leq d_D \leq 4.0 \text{ mm}$. In this range for use with geothermal brines or water as the continuous phase, the drops are nominally spherical with the possibility of being in the fluctuating ellipsoidal regime. For hexane, pentane or isobutane, experience at the University of Utah indicates that the drops are spherical. In the presence of impurities, such as occur in geothermal brines, there is further justification to presume this behavior. This is due to the fact that impurities tend to immobilize the interface. Thus, it is believed that little or no circulation will take place in drops of light hydrocarbons in water, or brine for diameters less than 4 mm. Further, in swarms, the terminal velocities of drops are decreased which makes it less probable that they are in the fluctuating ellipsoidal regime.

When the drops have low thermal conductivity, as is the case with hydrocarbons, it is likely that the governing resistance to heat transfer is internal to the drops. However, until the recent work of

Jacobs and Golafshani^(13,14), it was generally assumed that external resistances were governing. Letan^(9,10) and coworkers postulated models based purely on the hydrodynamics of the drops. Noting that at moderate Re_c , $Re_c > 500$, that the drops periodically shed their wakes, Letan, et al.⁽⁹⁾ postulated that the reduced heat transfer as compared to single drops was related to the shedding frequency and wake size. They argued that for long columns most of the heat transfer occurred in the wake shedding regime. Allowing for some empirical constants in the model, they were able to fit their own and some other data. Further, for long columns they could justify the use of a constant volumetric heat transfer coefficient. As their experiments were conducted at small temperature differences between the incoming fluids, they did not worry about temperature dependent fluid properties.

Following the lead of Letan and earlier investigators, Plass, Jacobs and Boehm⁽¹⁵⁾ ran a series of experiments to determine a volumetric heat transfer coefficient, U_v . They correlated both their own data and that of other investigators for organic fluids dispersed in water or geothermal brine. They claim an accuracy of $\pm 20\%$ for the following correlation

$$U_v = 1.2 \times 10^4 \epsilon \frac{\text{Btu}}{\text{hrft}^3 \text{ } ^\circ\text{F}} \quad (\text{for } \epsilon < 0.05) \quad (31)$$

$$U_v = [4.5 \times 10^4 (\epsilon - 0.05) e^{-0.57 G_D / G_c} + 600] \frac{\text{Btu}}{\text{hrft}^3 \text{ } ^\circ\text{F}} \quad (\text{for } \epsilon > 0.05) \quad (32)$$

where

$$U_v = \frac{\dot{Q}}{\text{Vol LMTD}} \quad (33)$$

The above equation was used in estimating the preheater length requirement for the 500 KW_e East Mesa combined boiler/preheater spray column.

For drops originally of 3.0 - 3.5 mm in diameter, Jacobs and Golafshani⁽¹³⁾ showed that the heat transfer is reasonably well-represented by Equations 31-32 when actual local holdup values are used. However in deriving Equations 31-32, Plass, et al.⁽¹⁵⁾ used the correlation of Johnson, et al.⁽¹⁶⁾ for holdup. The use of the latter correlation gave "sometimes agreement" with the data from East Mesa⁽¹⁰⁾. The degree of accuracy in predicting preheater length was approximately $\pm 20\%$ depending upon how the holdup was calculated. The calculated, detailed temperature profiles did not compare as well with the experiments. Jacobs and Golafshani⁽¹³⁾ also investigated a model using the assumption of no drop internal resistance to heat transfer and one where the heat transfer was governed by diffusion within the drop. This latter model showed better agreement, especially when it accounted for drop growth.

For drops less than 4.0 mm in diameter, it is thus recommended that final preheater spray column sizing be done using the conduction drop growth model of References 13 and 23. Preliminary sizing can be carried out using Equation 33.

For drops greater than 4 mm in diameter, it is highly probably that the drops will be in the fluctuating ellipsoidal regime. For this regime, both internal and external resistances to heat flow would have to be considered. For single drops, Sideman⁽¹⁷⁾ gives

$$Nu_d = 50 + 8.5 \times 10^{-3} Re_c Pr^{0.7} \quad (34)$$

for the external surface coefficient for oscillating drops for $150 < Re_c < 700$. A maximum deviation of 12% was reported. No reliable expression is available for $Re > 700$.

The internal resistance to heat transfer can be calculated by Equation 28 of Reference 14. It gives

$$Nu_d = 0.00375 \frac{Re_d Pr_d}{1 + \mu_d/\mu_c} \quad (35)$$

This equation is reported to be accurate to $\pm 20\%$.

It is not clear from the studies conducted to date whether increases in the surface heat transfer due to internal circulation will offset reduced surface area by going to larger drops. Before settling on a drop size for a given application, however, such a study is warranted.

Direct Contact Boiling Heat Transfer

The sizing of the boiler for the 500 KW_e unit at East Mesa was also done using an estimated value for the volumetric heat transfer coefficient. Based on all available data for light hydrocarbons and freons boiling in water, it was observed that

$$U_v = 48,000 \epsilon \frac{\text{Btu}}{\text{hr ft}^3 \text{ } ^\circ\text{F}} \quad (36)$$

where ϵ is estimated at the value just below the boiler in the preheater section. Although this yielded a reasonable estimate and appears to agree well with the 500 KW_e facilities operation, it has no basis in the physics of the boiling phenomenon. However, no correlation yet proposed does. Further, Walter⁽¹⁸⁾, in a recent Ph.D. dissertation at the University of Tennessee concludes, "There appears

to be no way to calculate a heat transfer coefficient". For the lack of anything better, Equation 36 is recommended.

DESIGN APPROACH FOR GEOTHERMAL APPLICATION

Based on laboratory and field experience with the 500 KW_e unit at East Mesa, it is clear that we can safely design a dcx for approach or pinch temperatures of 2.5°C (4.5°F). The pinch temperature is a necessary parameter in carrying out the thermodynamic analyses to select the optimum direct contact binary cycle for a given geothermal resource. It will set the flow rates of the two fluids, brine and working fluid once the working fluid is selected. The complete power system can be chosen utilizing the computer program DIRGEO (described in Reference 2 and prior reports developed under this contract)(19-22) or similar programs.

On the basis of the system thermodynamic analyses, the mass flow rates of brine and working fluid would be available for a given geothermal resource. The direct contactor pressure, working fluid boiling point, and the inlet and outlet temperatures for both the brine and working fluid would be established. One could now, utilizing this information, proceed to design the direct contactor.

The first thing that must be done is to decide on the drop size for the dispersed phase. Typically the nozzle diameter, d_n , will be from one half to two thirds as large as d_p . As the equation for the drop size, Equation 9, depends upon the critical jetting velocity as well as the critical jet diameter, which in turn depend on d_n , we must select d_n first. For light hydrocarbons in brine values of d_n less than 1.58 mm (1/16 inch) should result in drops from 3.0 to 3.2 mm

(~1/8 inch) in diameter. Such drops should remain nearly spherical with little internal circulation. A larger nozzle diameter will result in fluctuating drops whose behavior at high holdup could lead to an unstable column. This, of course, would need to be examined for the actual fluids selected in light of Figure 3.

After selecting d_n , the jetting velocity, V_j , can be calculated from Equation 1. This is the minimum velocity for the injection nozzles. Next the critical jetting velocity and jet diameter are determined from Equations 2-5. The critical jetting velocity, V_{jc} , cannot be exceeded. At the critical jetting velocity, the drops formed will be given by either Equation 6 or 7 depending upon the E_o number. One can operate the injector at any velocity between the jetting velocity and the critical jetting velocity at long as the Weber number is greater than two. This is necessary to prevent seeping.

Knowing the desired drop diameter, Equation 9 can be used to calculate the nozzle velocity, V_n . As long as the criteria mentioned above is satisfied, we will have selected the nozzle size. We can now proceed to determine the required number of perforations, or nozzles, in the dispersed phase distributor.

As the total mass flow rate of the disperse phase is known, as well as its temperature and pressure, we can calculate the volumetric

flow rate, $G_d = \frac{\dot{m}_d}{\rho_d}$. The number of nozzles required is

$$n = \frac{G_d}{\frac{\pi}{4} d_n^2 V_n} \quad (37)$$

The number of nozzles, of course, must be a whole number. We round off the calculation to the nearest one, making sure we do not cause a problem by exceeding the value of V_{jc} . Using the whole number, we calculate a new value of V_n from Equation 37. We then recalculate d_p from Equation 9. This is then the drop diameter.

Depending upon whether or not the drops are spherical or wobbling ellipsoidal, which can be checked roughly from Figure 3, we are ready to calculate the drop terminal velocity using either Equation 15 or 18, as is appropriate.

Next the flooding holdup, ϵ_f , should be calculated from Equation 27. The design value of holdup for the column should be selected as $0.9\epsilon_f$ or less as pointed out in Equation 28. Correcting for the amount of mass of geothermal brine that is vaporized in the boiler with the working fluid, the mass entering the preheater is determined prior to the actual calculation of ϵ_f . Knowing the exit conditions of the brine, the volumetric flow rate G_c is determined. We now know both G_d and G_c . Their ratio $G_d/G_c = R$. Using equation 24 and calculating Re_c for the drop we obtain m . ϵ_f is the small positive root of Equation 27.

Having established the holdup for the bottom of the preheater, we can now calculate the superficial velocity of the continuous phase at the bottom of the column above the dispersed phase injector. Either Equation 22 or 23 can be used. Equation 23 is sufficient unless the holdup is very high, i.e., > 0.35 . The superficial velocity is defined as the mean velocity of the continuous phase across the entire column. Thus,

$$D_C = \sqrt{\frac{4}{\pi} \frac{G_c}{(G_c/A)}} \quad (38)$$

gives the needed diameter of the column just above the dispersed phase injector.

If the column was isothermal, selecting the diameter just calculated would insure its holdup being constant along its entire length. With heat transfer, the holdup may vary. This is due to the fact that the density of the brine and selected working fluid can vary considerably with temperature if the column pressure is sufficiently high. If the working fluid density varies, the drop diameters will vary and thus their terminal velocity, etc. No where along the column length should the holdup exceed $0.9\epsilon_f$. Thus, we should next calculate the conditions at the top of the preheater in a manner similar to what was done at the bottom. The column diameter should be which ever is larger.

Having determined the column diameter, we should next check the overall size of the dispersed phase distribution plate. The nozzle holes should not be placed on a center to center distance less than $1.5 d_D$. With this type of layout, the overall distributor plate injector should not be larger in diameter than the column. This will insure that the drops can rise vertically.

The columns shown in Figures 1 and 2 both have a conical section in which the dispersed phase injector is located. References 2 and 6 indicate that the cone half angle should be about 16° . The distributor plate should be so located that the annular ring of open space around it should equal the cross-sectional area of the column.

This will insure that there is no undercarry of the dispersed phase and that there is good separation of the phases.

We are now ready to proceed with the calculation of the length of the preheater part of the column. If the thermodynamic properties of the working fluid liquid, and brine do not vary significantly with temperature, it is possible to calculate the length of the preheater from

$$L_{P.H.} = \frac{\dot{m}_d (h_{bp} - h_{in})_d}{U_v \left(\frac{\pi D_c^2}{4} \right) LMTD} \quad (39)$$

where h_{bp} is the enthalpy of the working fluid liquid at the boiling point, h_{in} is its enthalpy at the inlet, U_v is the volumetric heat transfer coefficient and LMTD is the log mean temperature difference across the preheater assuming counterflow. Unfortunately for fluids like isobutane, the specific heat varies considerably with temperature as was noted in the design of the 500 KW_e unit at East Mesa. Thus, it is necessary to evaluate the heat transfer in a number of steps along the preheater length. This was done by hand using Equations 31-33 for the 500 KW_e unit.⁽³⁾ For design studies it is recommended that the steady state computer program described in Reference 23 be used to determine the preheater length. The computer program is easily modified to include a variety of methods for estimating the heat transfer rate. For spherical drops, it is recommended that the variable diameter drop conduction model described in References 10 and 23 be used to determine the preheater length. For fluctuating ellipsoidal drops, Equations 34 and 35 can be used.

The length of the boiler section can be calculated using the value of U_v given in Equation 36. Due to the nature of the equation as discussed, it is not warranted to divide the boiler into segments. Thus the length of the boiler should be calculated as

$$L_b = \frac{\dot{m}(h_{\text{exit}} - h_{\text{b.p.}})_d}{\frac{\pi D_C^2}{4} U_v \text{LMTD}}$$

where LMTD is based on counterflow temperatures across the boiler.

Based on results of the 500 KW_e design, the length of the spray column should start at the top of the conical section housing the dispersed phase distributor.

It is recommended that the continuous phase injector or injectors be located in the middle of the boiler section to insure maximum heat transfer rates.

A disengagement section at the top of the column needs to be included as shown in Figure 2. This should be sufficiently large to insure no liquid carryover with the exiting vapor.

A sensitivity analysis should be made to determine possible input variations and the time constant for the column such as was done in Reference 23 for the 500 KW_e unit. The controls for the column should be designed based on this information and the resulting information indicating possible preheater and boiler length excursions.

Following the above procedures it is possible to design a highly reliable spray column for geothermal power applications. The techniques posed appear to be conservative based on experience with the 500 KW_e East Mesa unit. Further refinements will require

additional laboratory studies as mentioned in the discussions in this manual.

APPLICATIONS OF THE MANUAL TO ISOBUTANE-WATER SYSTEMS

Isobutane has been shown⁽²⁾ to be an excellent choice as a working fluid for geothermal brines when the brine temperature is above 300°F (~149°C). This section presents some calculations using the methods discussed in the manual to show the influence of various parameters on spray column design.

Establishing Drop Size

In designing a spray column, it is necessary to choose the drop size in order to establish the column hydrodynamics. Figure 4 was developed for the isobutane-H₂O system utilizing Equations 1-5. For a small nozzle diameter, it is clear that a wide range of nozzle velocities are possible between the working limits of V_j and V_{jc} . However, as the nozzle diameters increase, the range of permissible velocities decrease. Thus, it would appear safer to design using nozzles of < 3.0 mm in diameter if one wished to provide for significant variations in velocity. However, since the area of the nozzles is proportional to their diameter squared, near equivalent volume flow changes may occur for much smaller changes in velocity for the larger nozzles.

If one considers the actual drop sizes as a function of the nozzle velocities, it is clear from Figure 5 that nearly the same variation in drop diameters is possible over all nozzle sizes shown. However, the approximate nozzle diameter limit for non-circulating drops is less than 2.5 mm (7/64 inch) and at velocities near the

jetting velocity. As it is most easy to hydrodynamically design spray columns for situations where the drops behave as rigid spheres, typical nozzle diameters of 1.5 - 2.0 mm are normally chosen. Plugging or partial plugging of some holes can lead to velocities exceeding the critical jetting velocity. Reductions in working fluid flow rate can cause drop diameters to exceed the limit for rigid sphere behavior. Thus, fluctuations in spray column operation can occur even without recirculation. Nonetheless, the advantages of spray column direct contact heat exchange make them attractive to pursue. It should also be noted that the limits for rigid drop behavior and critical jetting velocity are experimentally established and are, in general, conservative.

Figure 6 shows the terminal velocities calculated using Equation 15 for rigid drops and Equation 18 for fluctuating drops. For all drops formed between the limits on Weber number and V_{jc} for the nozzle dimensions considered, isobutane drops in H_2O fall within either the spherical or fluctuating ellipsoidal regime shown in Figure 3 as $\log M$ is approximately -12.16 and $.7 \leq Eo \leq 2.6$. It should further be noted that the terminal velocity is higher for drops with internal circulation by up to 50%. Thus, it should be possible to increase the throughput by increasing drop size. However, as the path of larger drops is not vertical, coalescence is more likely to occur. Therefore, at this time, it does not appear prudent to significantly increase drop size without further experimental data on drop hydrodynamics, especially that for drop swarms.

Establishing Flooding Limits

Data on flooding in spray columns has primarily been determined in small diameter columns where change of phase of the volatile fluid does not occur. This has been noted in the preceding sections, and is the case for Equation 27. This correlation was developed for the spherical drop regime. Equation 23 presents a correlation for the continuous phase superficial velocity also for the rigid sphere regime. Utilizing these two correlations, the nondimensional, continuous phase, superficial velocity as a function of holdup, ϵ , for a range of the volumetric flow ratio, R , is shown in Figure 7. Note the flooding limit, ϵ_f , and $0.9\epsilon_f$ and $0.8\epsilon_f$ are also shown. The range of R is consistent with the use of dchx systems with moderate temperature geothermal brines and solar ponds. For this range, the flooding condition corresponds to a nearly maximum value of continuous phase superficial velocity. The peak value of ϵ_f varies from nearly 0.34 at an R of 2.25 to 0.25 for $R = 0.5$. In each case, the superficial velocity only undergoes a small change in order to reduce to $0.8\epsilon_f$ or $0.9\epsilon_f$.

The advantage for heat transfer of operating close to the flooding point is shown in Figure 8. For the case of small values of R , as is the case for solar ponds, the correlation of Reference 15 indicates volumetric heat transfer coefficients of nearly 8,000 Btu/hrft³ °F (148.8 Kw/m³°K) when operated near flooding. Of course, it should be noted that in the liquid preheating regime, such a low value of R leads to low utilization of the heating capacity of the continuous phase brine. In typical geothermal applications, it is desired to utilize a considerable portion of the thermal capacity of

the hot liquid brine. With an organic working fluid such as isobutane, a typical value for R would be around two. The correlation of Reference 15 indicates a decrease in U_v with increasing R , but an increase with ϵ . Since the rate of increase of U_v with increasing ϵ is less for higher values of R , operating further away from the flooding point does not, as significantly, affect U_v as it would at lower R , more typical of solar pond applications.

Figure 8 can be used for preliminary design applications provided we keep in mind the data from which it was determined. The limitations are: (a) the dispersed phase is an organic fluid; (b) the continuous phase is brine or water; (c) the drops are less than 5.0 mm in diameter; and (d) the correlation is good to only within $\pm 20\%$.

EXAMPLE OF DCHX DESIGN

Consider the design of a dchx preheater-boiler for brine entering at a rate of 85,840 lbm/hr. Isobutane at a flow rate of 91,434 lb/hr enters the dchx at 86°F. These mass flow rates correspond to 327.9 gpm of IC_4 and 192 gpm of brine. The boiling point of the IC_4 is 244.5°F (corresponding to a vapor pressure of 400 psi), and the temperature of the brine entering the preheater is 251°F. The brine exits the column at 129.5°F.

The change in enthalpy of the isobutane across the preheater is 10.332×10^6 Btu/hr. In order to size the column, we must first choose the size drops we desire. Let us choose 3.5 mm diameter at entrance. This size can be achieved for nozzle diameters less than 1.95 mm (see Figure 5). The maximum velocity, V_{jc} , will be 540 mm/sec

(1.772 ft/sec) for this size nozzle. Let us choose a smaller nozzle diameter. One-sixteenth inch diameter holes can be drilled (1.587 mm) to form nozzles. For this size nozzle, the critical jetting velocity is 630 mm/sec and the critical drop diameter $d_{DC} = 2.5$ mm. Using Equations 2 and 9, we can solve for d_{jc} and V_n , the nozzle velocity. The equations yield $d_{jc} = 1.46$ mm and $V_n = 430$ mm/sec (1.47 ft/sec). This velocity is in the midpoint range between the V_{jc} and $W_e = 2.0$ limits. The number of holes required for the distributor plate can be calculated from Equation 37. The nearest whole number yields 23,328 nozzles. The area of the nozzles is 0.497 ft² or 0.0462 m².

Next we must calculate the terminal velocity of the drops. As they are in the spherical regime (see Figure 6), the terminal velocity can be calculated from Equation 15. The terminal velocity at the bottom of the column is 280 mm/sec or 0.919 ft/sec. Before determining the superficial velocity of either phase, we must next calculate the holdup at flooding.

Using the terminal velocity, the Reynolds number of a drop is calculated. This is necessary to establish the Re_c regime. As the Re_c is greater than 500, $Re_c = 2212$, we obtain $m = 1.39$. From Equation 27, we next find the value of the holdup at flooding, ϵ_f . Given the ratio of flow rates $R = 1.708$ and $m = 1.39$, ϵ_f is 0.326. If we choose to design for $0.9\epsilon_f$, then $\epsilon = 0.2934$. From Equation 23, we can calculate the superficial velocity at the bottom of the column. The chosen conditions yield $G_c/A = 0.0784$ ft/sec or 23.88 mm/sec.

Since we know the volumetric flow rate of the brine G_c , the diameter of the column proper at its base can be calculated, from

Equation 38, to be 2.80 ft or 0.855 m. Although this would be an appropriate diameter for the dchx column for conditions near the entry of the working fluid, large changes in the density of isobutane can occur at pressures near 400 psi as it is heated to its saturation temperature. Thus, the flooding conditions will have to be checked at several locations along the column length in the preheater section. At the top of the preheater section, the 3.5 mm drops will have grown to a diameter of nearly 4.0 mm. The terminal velocity will increase and generally the deviation in holdup will be to one further from the flooding point.

In order to establish the approximate length of the preheater, Equations 32 and 39 can be used. From Equation 32, U_v is approximately 4,650 Btu/hrft³ °F or 86.5 Kw/m³°K. The LMTD for the preheater is 21.8°F. Thus, from Equation 39, the approximate length of the preheater is 16.55 ft or 5.04 m.

In the boiler section, 5.85×10^6 Btu/hr will be transferred to the isobutane. Equation 36 yields a $U_v = 14,083$ Btu/hrft³ °F. The LMTD across the boiler is 31.5°F. Thus, the boiler volume is 13.19 ft³. The boiler length is 2.14 ft or 0.65 m.

We now have a preliminary sizing of the column proper. The diameter is 2.8 ft (0.855 m) and the combined length of the preheater and boiler is 18.7 ft (5.70 m). Using these preliminary dimensions, the steady state computer program described in Reference 23 can be used to refine the size for the preheater. Such a calculation yields an overall length of 19 ft (5.8 m).

For a vapor-mist disengagement section, a height of two diameters is suggested (5.6 ft or 1.71 m). Two rows of chevron mist eliminators

located mid-way in the disengagement section would eliminate any droplet carry-over. The chevrons should be two inches high and inclined at 60° from the vertical.

The brine injector would be located near the midpoint of the boiler section. An injector might be designed to distribute the brine horizontally from a single tube. It should yield a horizontal velocity no greater than G_c/A at the wall of the column. This would require the injector distribution over a height of 7.4 inches or 18.9 cm. A design such as used for the 500 KW_e East Mesa DCHX unit described in Reference 3, would be satisfactory.

The isobutane distribution plate, as noted, would require 23,328 1/16 inch diameter holes spaced over 6.16 ft². This is equivalent to one hole every 0.038 square inches, or 0.22 inches or 5.59 mm between centers. The distributor should be located in the conical frustrum section below the column proper where the diameter is 3.96 ft. Ideally, the frustrum 1/2 angle would be 15° . This will allow for the brine to pass by the IC₄ distributor with no increase in the continuous phase superficial velocity. The brine exit would be located on the pressure head below the column in a manner similar to that shown on the schematic of Figure 1.

This design example used flow conditions identical to those for the East Mesa DCHX observed on November 12, 1980⁽³⁾. The East Mesa DCHX had a diameter of 3.67 ft. Under these flow conditions the holdup, ϵ , for the East Mesa DCHX was only 0.227, which was only 70% of the flooding value. The resulting U_v calculated from Equation 32 would have been 2,718 Btu/hrft³ °F. The experiment indicated a U_v of

2,390 Btu/hrft³ °F. The comparison is within the $\pm 20\%$ claimed by Reference 15.

NOMENCLATURE

A	Internal cross-sectional area of the column
d_D	Drop diameter
d_{DC}	Critical drop size according to Treybal
d_{jc}	Drop diameter at critical jetting velocity
d_n	Nozzle diameter
$\Delta\rho$	$ (\rho_c - \rho_d) $
D_C	Column diameter
E_c, E_d	Dispersion coefficients, empirically determined constants used in one dimensional conservation equations to account for backmixing
E_o	Eotvos number, defined by Equation 8 and 11
ϵ	Void fraction, local global fraction of volume occupied by the dispersed phase
ϵ_f	Void fraction at the flooding point
f	Drag coefficient defined by Equation 16 for rigid spheres and Equation 17 for surface mobile spheres
g	Gravity
G_c	Volumetric flow rate of continuous phase
G_c/A	Superficial velocity of the continuous phase
G_D	Volumetric flow rate of dispersed phase
h_{bp}	Enthalpy of liquid at the boiling point
h_{exit}	Enthalpy at exit from the column

h_{in}	Inlet value of enthalpy
H	Defined by Equation 21
J	Defined by Equation 19 and 20
k, n	Constants in Equation 22
K	Defined by Equation 4
K_1	Defined by Equation 13
L	Length of column
LMTD	Log mean temperature difference for countercurrent flow
$L_{p.H.}$	Preheater length
m	Constant in Equation 23 defined by Equation 24
μ_c	Dynamic viscosity of continuous phase
M-group	Defined by Equation 12
n	In Equation 37, the number of nozzles required in the dispersion plate
Nu_d	Nusselt Number defined on the basis of drop diameter
Pr	Prandtl number
$\frac{Q}{Vol}$	Total local heat transfer between phases per unit volume
ρ_c	Continuous phase density
ρ_d	Dispersed phase density
R	G_D/G_C
Re	Reynolds number, defined by Equation 10
σ	Interfacial tension
U_v	Volumetric heat transfer coefficient defined by Equation 33

V_{cj}	Maximum velocity in nozzles to insure near uniform drop formation
V_j	Jetting velocity, minimum velocity in nozzle to insure all nozzles are flowing
V_n	Actual nozzle velocity corresponding to d_n , actual nozzle diameter and flow rate of dispersed phase
V_T	Terminal drop velocity for surface mobile spheres this is calculated from Equation 18
W_e	Weber number, $\frac{V_n^2 d_n^2 \rho_d}{\sigma}$

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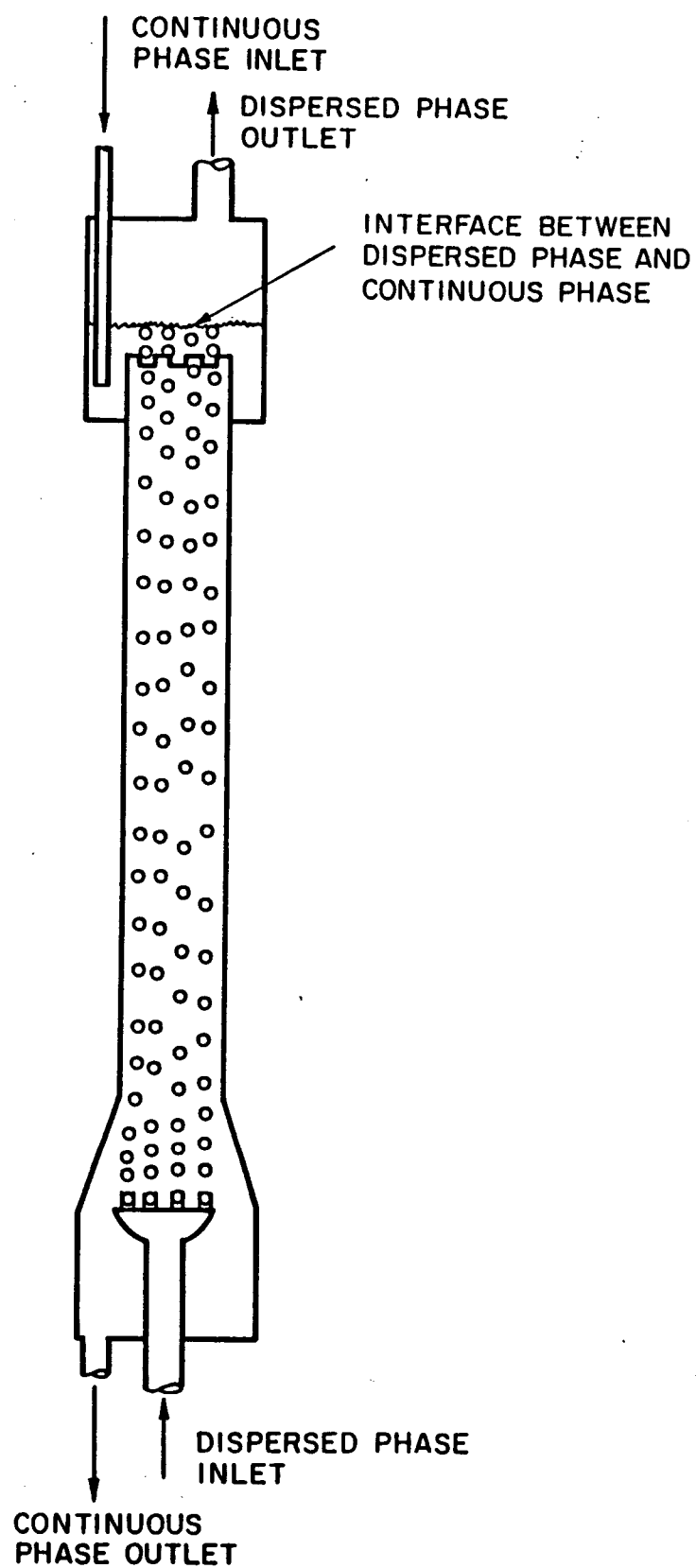


Figure 1. Direct Contact Spray Tower for Liquid-Liquid Heat Exchange.

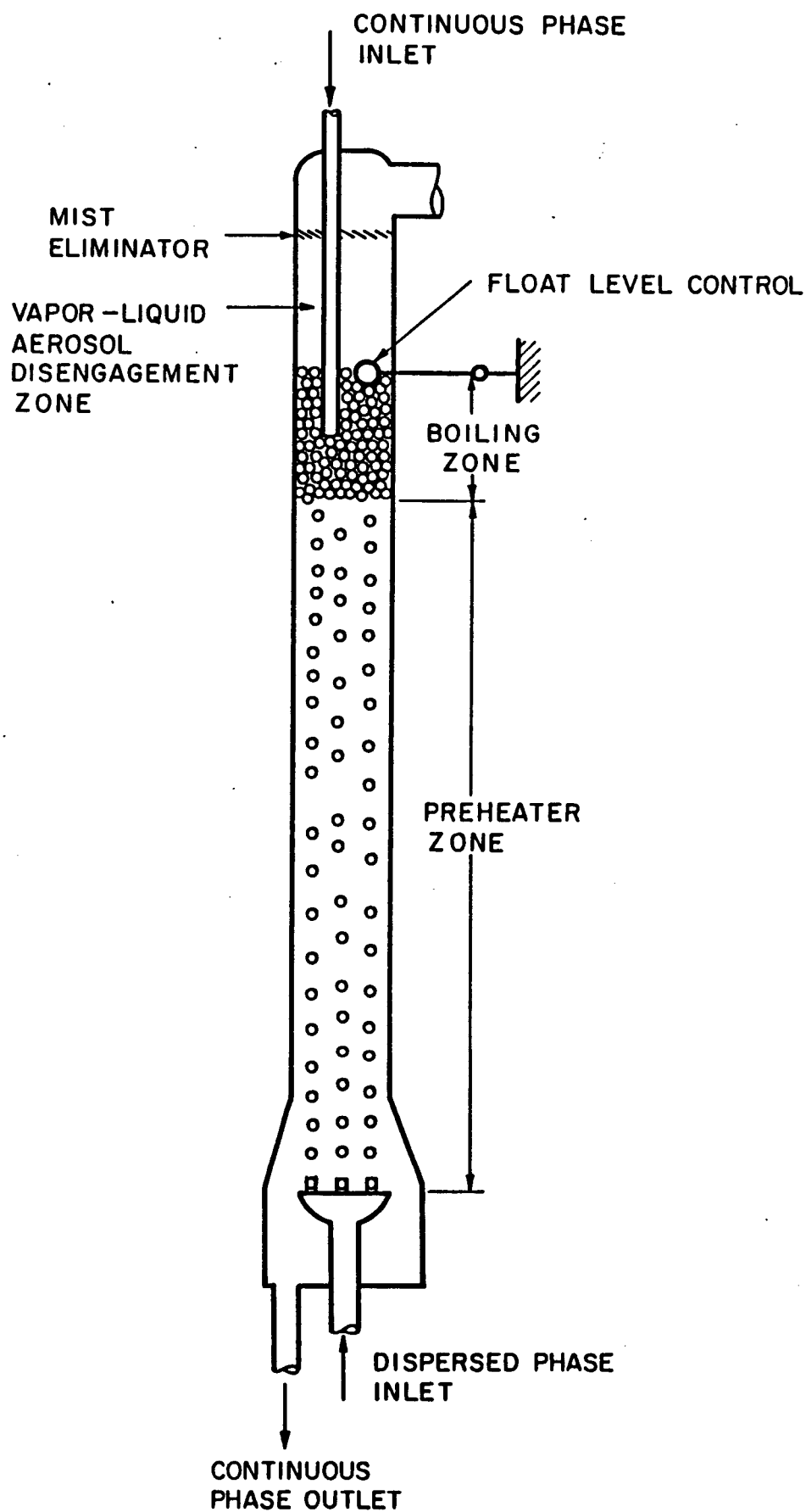


Figure 2. Direct Contact Spray Tower for Preheating and Boiling Dispersed Phase.

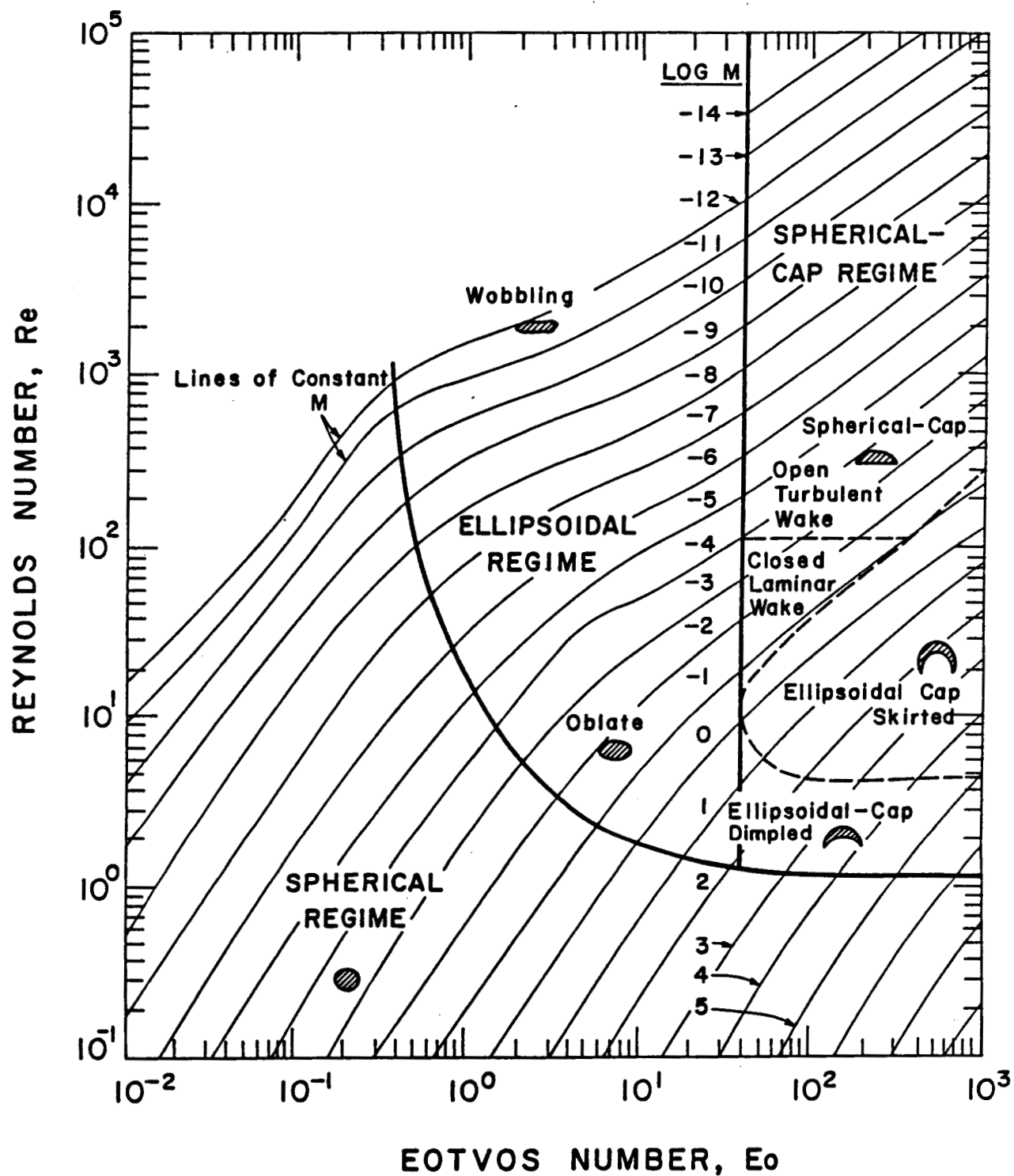


Figure 3. Drop Characterization Map.

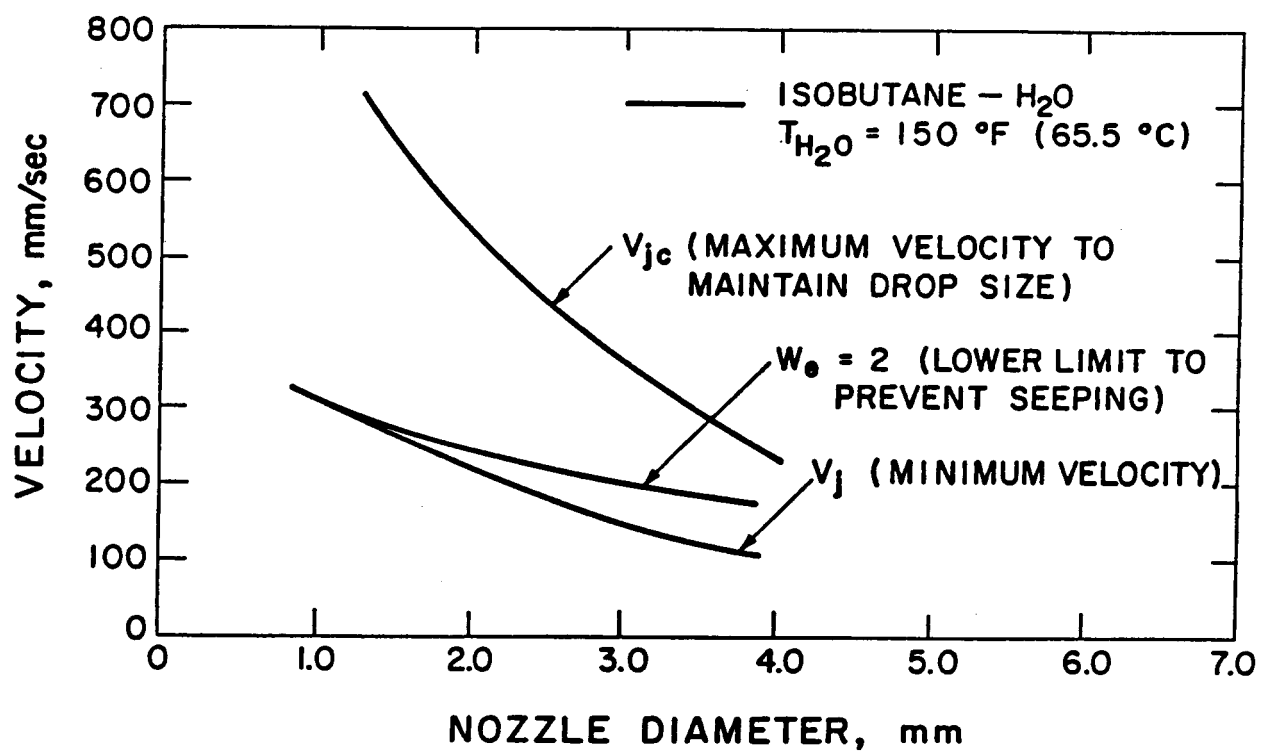


Figure 4. Typical Limiting Velocities for Changes in Nozzle Diameter.

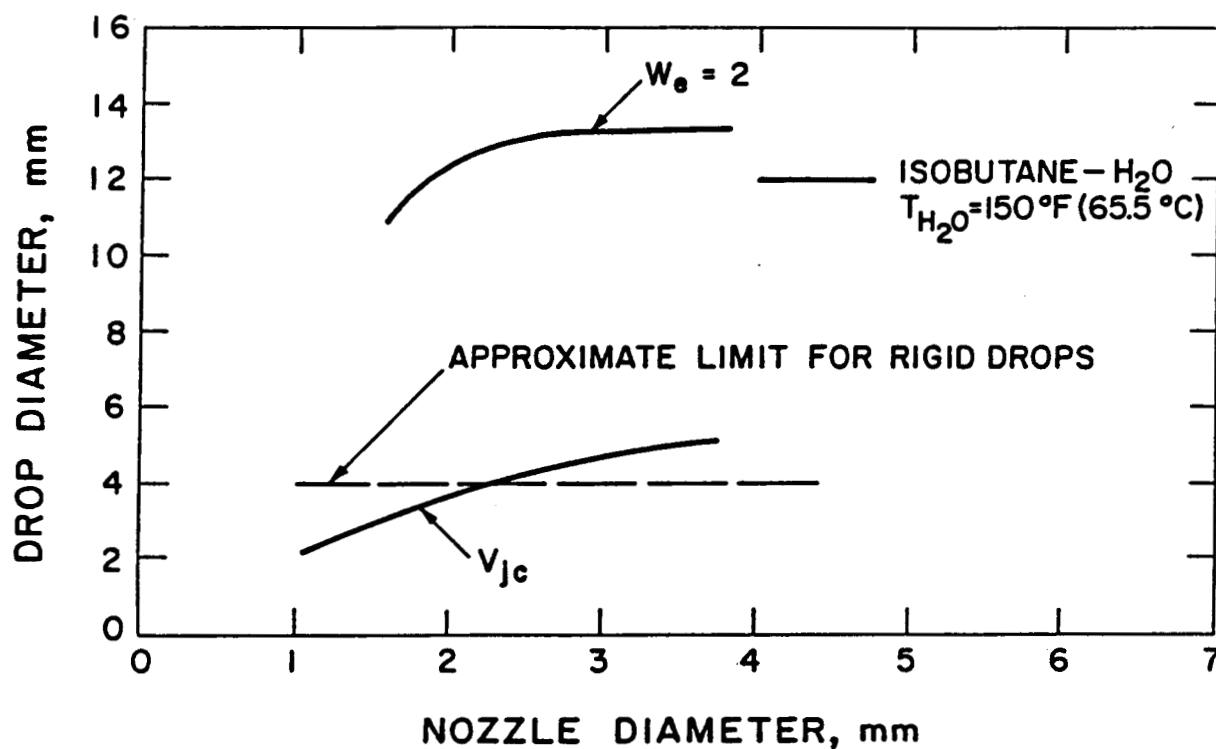


Figure 5. Variation in Drop Sizes Produced by Nozzles Within Operating Velocity Limits.

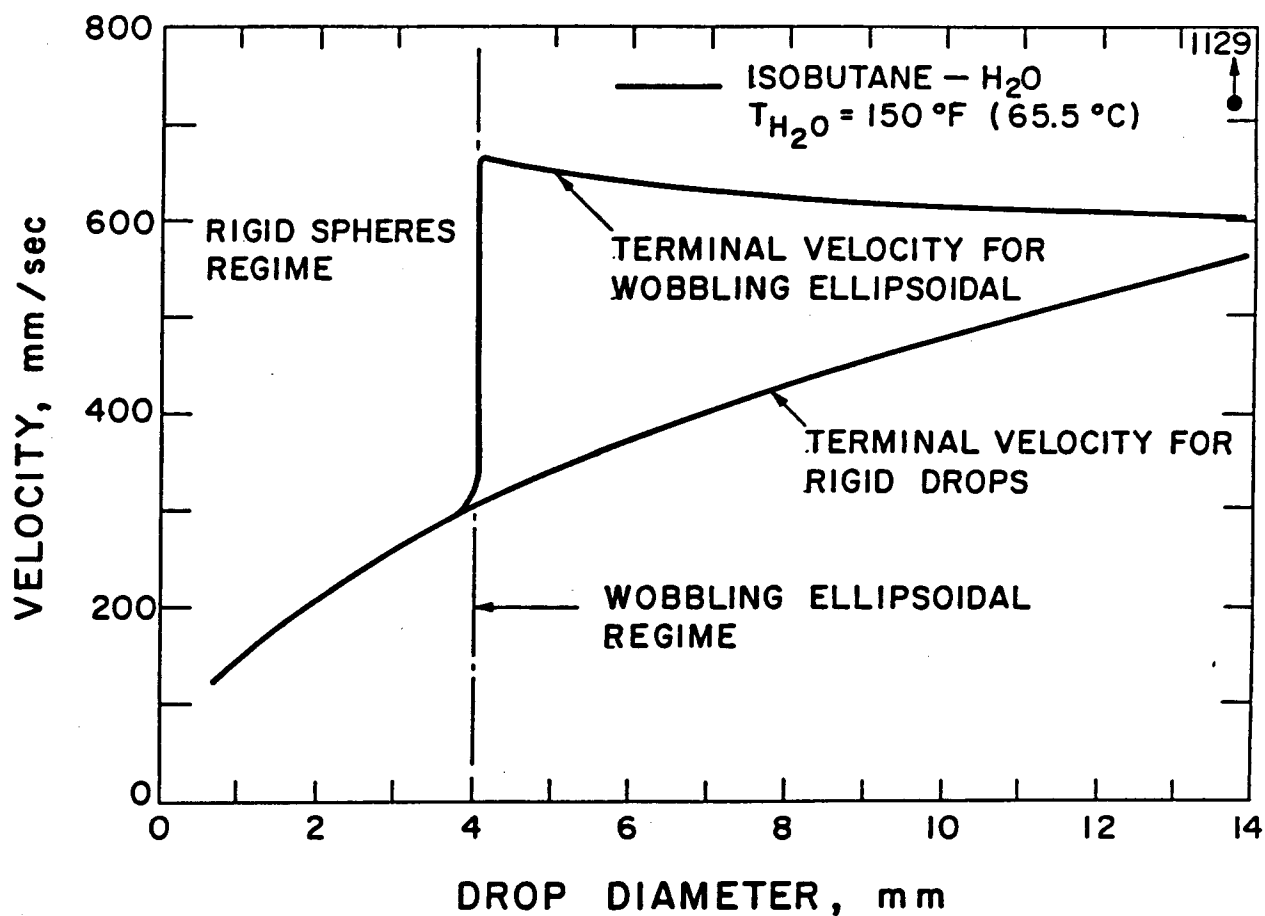


Figure 6. Terminal Velocities for Rigid Spheres and Fluctuating Drops According to Figure 3.

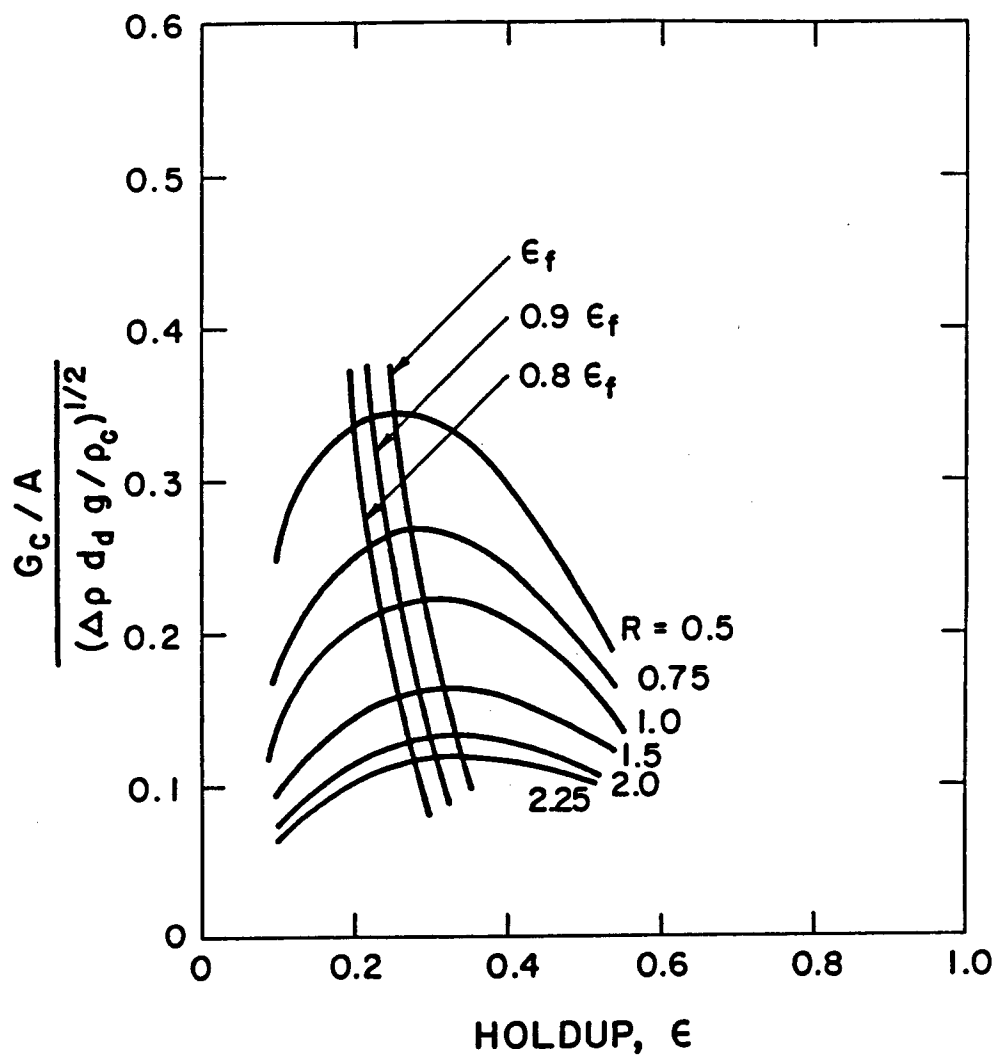


Figure 7. Superficial Continuous Phase Velocity as a Function of Holdup for Spherical Drops.

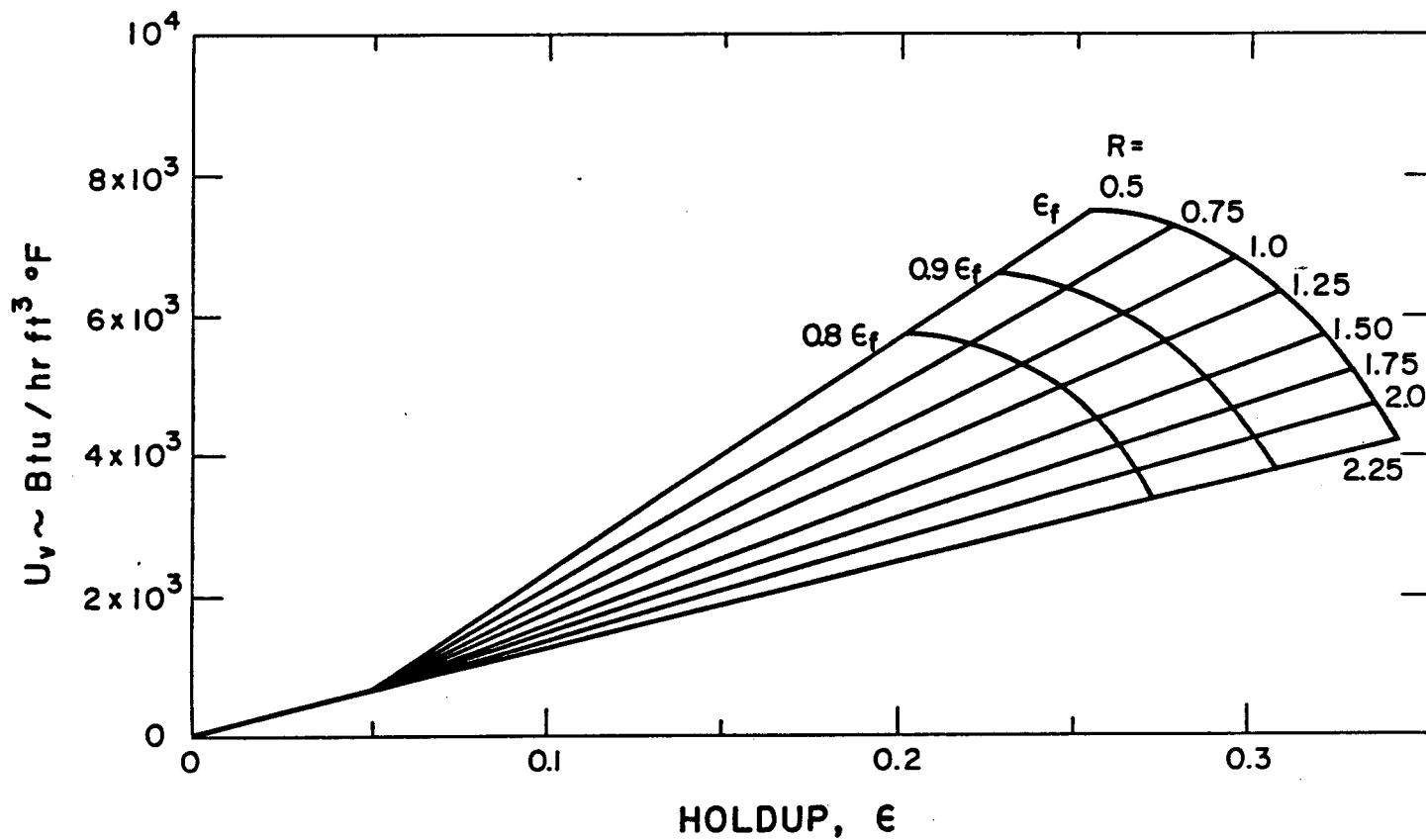


Figure 8. Volumetric Heat Transfer Coefficient as a Function of Holdup from Reference 15.