Heat Transfer in Liquid-Liquid Spray Towers

E. J. GRESKOVICH
Esso Research and Engineering Company, Florham Park, New Jersey
PAUL BARTON and R. E. HERSH
Pennsylvania State University, University Park, Pennsylvania

An experimental investigation of heat transfer rates in which were used three spray towers varying from 4 to 6 in. in diameter and 7 to 10 ft. in height, and distributor elements with hole sizes varying from 1/8 to 1/64 in. in diameter was carried out. The immiscible liquid system consisted of water and kerosene (normal boiling point range 395° to 490°F.). Volumetric heat transfer coefficients ranged from 100 to 3,600 B.t.u./(hr.)(cu. ft.)(°F.) and the overall film heat transfer coefficients ranged from 4 to 23 B.t.u./(hr.)(sq. ft.)(°F.). Height equivalent to a theoretical stage varied from 1 to 8 ft. Scale-up correlations relating these rate data to operating variables were developed.

There has been extensive development of more efficient heat exchangers for special purposes in the past decade. One area receiving much attention is that of direct-contact, liquid-liquid heat exchange. Even though this method of transferring heat has been practiced for many years, such as direct-contact cooling in propane dewaxing of lube oils as described by Bray et al. (*2*), recent interest and research in the area of seawater desalination have stimulated the investigation of scale-up of liquid-liquid spray towers for heat exchange.

The advantages of direct-contact, liquid-liquid heat exchange over other methods for liquid-liquid heat exchange are:

1. Heat transfer rates are not reduced by corrosion and scaling.
2. Large amounts of corrosion-resistant metals are not needed to provide area for heat transfer.
3. Heat flux can be maintained high at temperature differentials as low as 1° to 2°F., which is generally not the case for shell and tube heat exchangers.

The disadvantages of the direct-contact system are:

1. Large equipment diameter and fluid inventory are necessary.
2. For countercurrent operations, the relative velocity of the fluids is limited by density difference. Thus the heat transfer rate cannot be increased by increasing turbulence.
3. The application of direct-contact heat transfer is limited, since for a given fluid to be heated or cooled, the contacting fluid should, among other things, be practically immiscible.

Fundamental investigations involving direct-contact, liquid-liquid heat transfer have appeared sporadically in the literature; however those investigations concerning droplets or swarms of droplets in spray towers are of immediate interest. Correlations of observed heat transfer data with theoretical models such as those of Sideman and Shabtai (*13*) and Elzinga and Banchero (*3*) have been presented for liquid drops. Collection and correlation of heat transfer data in liquid-liquid spray towers both on small- and pilot-plant scales have been carried out by investigators such as Garwin and Smith (*5*), Pierce et al. (*11*), Johnson et al. (9), Woodward (*17*), Bauerle and Ahlert (*1*), and Sukhatme and Hurwitz (*15*). Sagar and co-workers (*12*), in addition, have studied heat transfer in a perforated plate tower. Pilot-plant studies have been undertaken by the Food Machinery and Chemical Corporation (*4*) in an attempt to scale up liquid-liquid spray towers for heat transfer in a saline water conversion process.

End effects in spray tower heat exchangers have been discussed by Letan and Kehat (*10*). The most complete treatment of direct contact heat transfer between immiscible liquids is that of Sideman (*14*).

Various investigations have been carried out utilizing direct-contact heat exchange in equipment other than spray towers. Grover and Knudsen (7) investigated heat transfer in turbulent cocurrent flow of a petroleum solvent and water. Wilke et al. (*16*) have studied direct-contact seawater evaporation using a chlorinated organic solvent.

In view of the wide interest in liquid-liquid heat transfer, especially for saline water conversion and liquid extraction processes, there is need for additional studies describing the effects of operating variables and obtaining scale-up correlations. It is the object of this presentation to assist in establishing scale-up correlations.

THEORY

Heat Transfer Coefficients

In discussing heat exchange between a liquid drop in the presence of an immiscible liquid medium, one can express the rate of heat transfer as

$$q = UA(\Delta T)_m \qquad (1)$$

The mean overall heat transfer coefficient can be expressed as

$$\frac{1}{UA} = \frac{1}{h_o A} + \frac{1}{h_w A} \tag{2}$$

The total area available for heat transfer (total interfacial area) can be calculated with the average drop diameter:

$$A = \pi n d^2 \tag{3}$$

It would be more exact to obtain a drop size distribution throughout the exchanger and thus obtain a more accurate measure of the total interfacial area; however, in many cases, properly designed distributors can produce reasonably uniform drops, and average drop size measurements are within reasonable experimental error.

The holdup of the dispersed phase $(1 - \varepsilon)$ is defined as the fraction of the effective tower volume occupied by the dispersed phase; then

$$1 - \varepsilon = \tfrac{1}{6} \cdot n \pi d^3 / V \tag{4}$$

Rearranging and combining Equations (3) and (4), one obtains the following:

$$A = 6(1 - \varepsilon)\, V/d \tag{5}$$

Therefore the average drop diameter and the dispersed phase holdup can be measured to yield the total interfacial area available for heat transfer.

For systems in which most of the resistance to heat transfer is in one phase, the individual heat transfer coefficient for that phase approaches the value of the overall coefficient. For example, as the size of the dispersed phase droplets decreases, the degree of internal circulation within the droplets decreases approaching the stagnant drop model. Physical properties, mainly viscosity, can also influence the internal circulation. Scale-up correlations based on dimensionless correlating parameters such as Reynolds, Nusselt, and Prandtl numbers use these values of U and the physical properties of the controlling phase. For systems with significant resistances in both phases, it becomes necessary to estimate the individual coefficients for correlating purposes.

The Nusselt number for the controlling phase can be defined as

$$N_{Nu} = \frac{h_c d}{k_c} \cong \frac{Ud}{k_c} \tag{6}$$

The Reynolds number of the controlling phase is

$$N_{Re} = \frac{d v_r \rho_c}{\mu_c} \tag{7}$$

where

$$v_r = \frac{v_{so}}{1 - \varepsilon} + \frac{v_{sw}}{\varepsilon} \quad \text{for oil dispersed} \tag{8a}$$

or

$$v_r = \frac{v_{sw}}{1 - \varepsilon} + \frac{v_{so}}{\varepsilon} \quad \text{for water dispersed} \tag{8b}$$

Since only a single system over a relatively small temperature range is employed in this study, the Prandtl number is not included as a correlating parameter. Although the presence of the Prandtl number in a final overall correlation is important, it was found necessary to keep the physical properties (C_p, u, k) constant in these investigations to gain better insight into the interaction of the hydrodynamic variables and the heat transfer mechanisms.

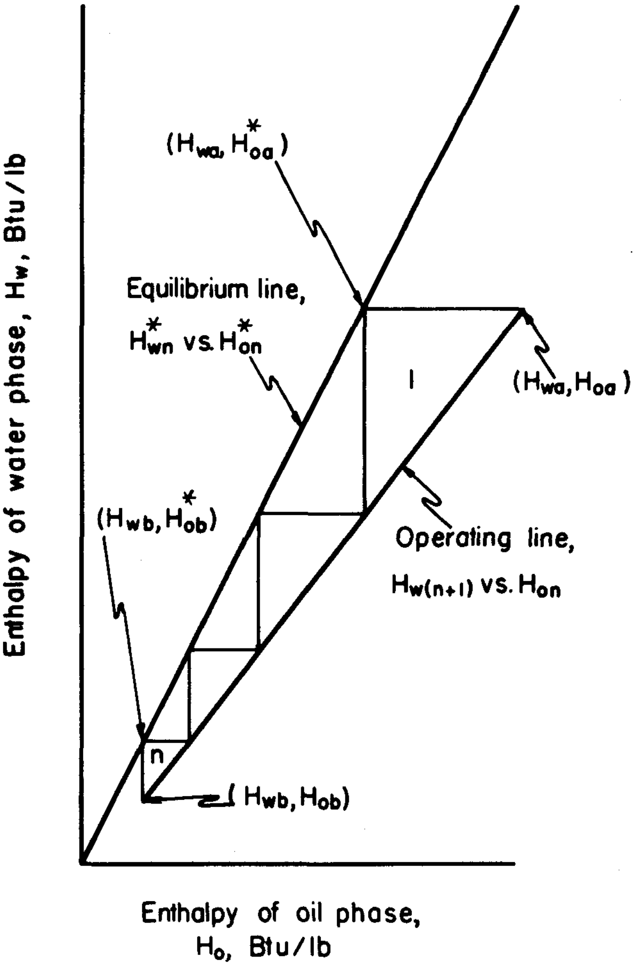

Fig. 1. Graphical determination of the number of theoretical stages (N_p) in countercurrent heat transfer.

For some systems it is not possible to determine the average drop size because the continuous phase is opaque. Since interfacial areas cannot then be measured to determine film heat transfer coefficients, it becomes necessary to use a volumetric heat transfer coefficient defined as follows:

$$(Ua) = \frac{q}{V(\Delta T)_m} \tag{9}$$

From Equation (9) it can be noted that the volumetric coefficient may not entirely take into account such variables as drop size, relative flow rates of the phases, and dispersed phase holdup. However, if these hydrodynamic variables can be kept constant, the volumetric coefficient may be the most effective, readily available variable for use in scale-up correlations.

THEORETICAL STAGES AND TRANSFER UNITS

The number of theoretical stages or theoretical plates required for a given countercurrent heat exchange duty can be obtained graphically as shown in Figure 1. The majority of the previous publications utilize the transfer unit concept, whereas in this work the height equivalent to a theoretical stage (H_p) concept will be discussed. The usual simplifying assumptions hold for construction of straight operating lines.

The number of theoretical stages can be calculated algebraically if one knows the terminal enthalpies and providing that both the operating and equilibrium lines are linear, using equations of the type summarized by Hartland (*8*). The criteria for a straight equilibrium line are that only sensible heat is transferred and that the ratio of specific heats of the phases be constant throughout the exchanger. For the case represented in Figure 1, the number of theoretical stages is given by

$$N_p = \frac{\ln \dfrac{H_{oa} - H^*_{oa}}{H_{ob} - H^*_{ob}}}{\ln \dfrac{H_{oa} - H_{ob}}{H^*_{oa} - H^*_{ob}}} = \frac{\ln \dfrac{\Delta T_a}{\Delta T_b}}{\ln \dfrac{w_w c_{pw}}{w_o c_{po}}} \tag{10}$$

For the case of constant driving force throughout the exchanger brought about by matching $w_o c_{po}$ and $w_w c_{pw}$, that is, setting $w_o c_{po} = w_w c_{pw}$, the operating line and equilibrium line are parallel and the number of theoretical stages is calculated by

$$N_p = \frac{H_{oa} - H_{ob}}{H_{oa} - H^*_{oa}} = \frac{T_{oa} - T_{ob}}{\Delta t_a} \tag{11}$$

The number of overall heat transfer units based on the oil phase required for the countercurrent heat exchange duty shown in Figure 1 is calculated by

$$N_{tOo} = \frac{\ln \dfrac{H_{oa} - H^*_{oa}}{H_{ob} - H^*_{ob}}}{1 - \dfrac{H^*_{oa} - H^*_{ob}}{H_{oa} - H_{ob}}} = \frac{\ln \dfrac{\Delta T_a}{\Delta T_b}}{1 - \dfrac{w_o c_{po}}{w_w c_{pw}}} \tag{12}$$

The number of transfer units based on the water phase can be obtained by

$$N_{tOw} = \frac{w_o c_{po}}{w_w c_{pw}} N_{tOo} \tag{13}$$

The relationship between the number of transfer units based on the oil phase and the number of theoretical stages for nonparallel operating and equilibrium lines is given by

$$\frac{N_p}{N_{tOo}} = \frac{1 - \dfrac{w_o c_{po}}{w_w c_{pw}}}{\ln \dfrac{w_w c_{pw}}{w_o c_{po}}} \tag{14}$$

For parallel operating and equilibrium lines, the number of transfer units equals the number of theoretical stages.

Geoffrin (*6*) has shown that the heat transfer requirement per theoretical stage, as expressed by the product of the overall film coefficient and the interfacial area, is theoretically independent of the driving forces over a temperature range in which the change in U with temperature is negligible. It is related to flow rates and heat capacities as follows:

$$(UA)_{\text{theoretical stage}} = \frac{w_o c_{po}}{1 - \dfrac{w_o c_{po}}{w_w c_{pw}}} \ln \frac{w_w c_{pw}}{w_o c_{po}} \tag{15}$$

For the case of constant driving force throughout the exchange the heat transfer requirement per theoretical stage is given by

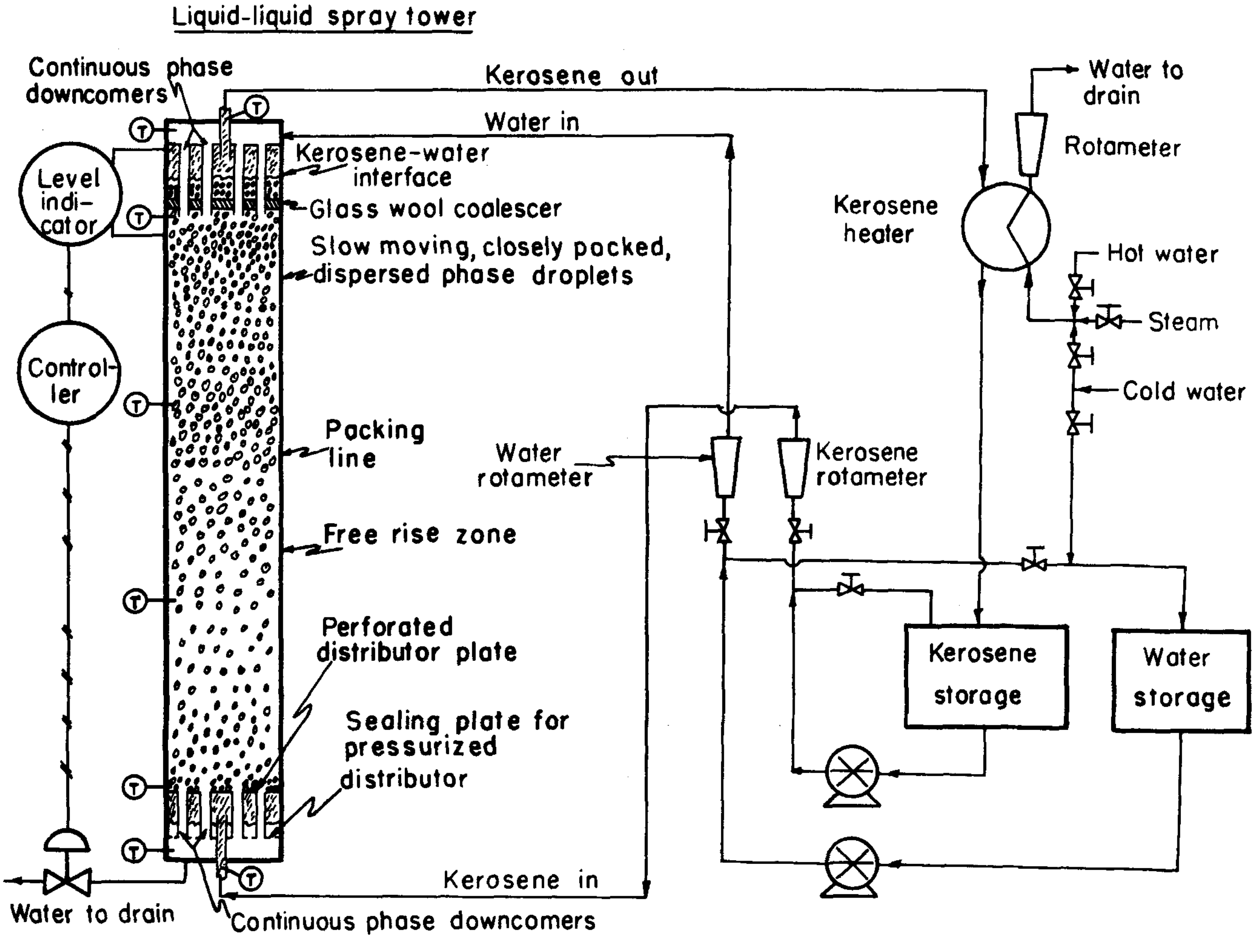

Fig. 2. Direct-contact, liquid-liquid heat transfer apparatus.

TABLE 1. COLUMN DIMENSIONS

Column designation	Inside diameter, in.	Effective height, ft.
A	6	10
B	4	10
C	6	7

$$(UA)_{\text{theoretical stage}} = w_w c_{pw} = w_o c_{po} \tag{16}$$

Expression of experimental values of (*UA*) per theoretical stage as a function of wc_p with other system variables as parameters will indicate which of the variables are important for scale-up and whether driving force in actuality can be neglected as a variable.

The transfer requirement per transfer unit based on the oil phase is found by substituting Equations (14) and (15) into

$$N_{tOo}\,(UA)_{\text{transfer unit}} = N_p\,(UA)_{\text{theoretical stage}} \tag{17}$$

to yield

$$(UA)_{\text{transfer unit}} = w_o c_{po} = w_w c_{pw} \tag{18}$$

Equation (18) shows that the heat transfer requirements per overall transfer unit are proportional to the throughput of the phase on which the transfer unit is based regardless of the wc_p ratio and that the requirements can be correlated independently of the driving forces with the constraint that *U* is not appreciably affected by temperature in the range under consideration.

It would appear that correlations or data reporting based on the values of *UA* requirements per transfer unit or theoretical stage divided by the appropriate wc_p expression [Equations (15), (16), (18)] might be more productive than the use of equivalent heights per se.

APPARATUS AND PROCEDURES

The apparatus, shown in Figure 2, was operated as a conventional spray tower, and since such an operation has been well described in previous literature, only a limited discussion will be presented here.

Three vertical columns constructed of flanged glass pipe were used in this study as presented in Table 1. These columns were fitted with various distributor elements to disperse one of the liquid phases. These elements, listed in Table 2, were fabricated from $\frac{3}{8}$-in. thick brass plates drilled with holes on concentric circles. Distributors *A*, *B*, and *E* were operated hydraulically free; that is, the liquid to be dispersed was forced through the holes in the distributor plate by density difference. Distributors *C* and *D* were sealed and pressurized by the pump so that the liquid to be dispersed was forced through the holes in the distributor plate at a faster rate than possible by density difference.

Tubular downcomers through the distributors and the coalescing zone facilitated uninterrupted flow of the continuous phase.

The position of the interface between the continuous phase and the recoalesced dispersed phase was maintained by a liquid level controller which operated the exit valve for the continuous phase. For kerosene dispersed studies, the interface was maintained at the top of the column. For water dispersed studies, the distributor was placed at the top of the column, and the interface was maintained at the bottom of the column; that is, the top and bottom parts of the column shown in Figure 1 were exchanged.

The column and auxiliary equipment were lagged with 1-in. thick fiber glass insulation. The temperature of the inlet water to the column was usually 50° to 55°F. and the temperature of the kerosene feed was typically 80° to 140°F. The kerosene used had a normal boiling point range of 395° to 490°F., a specific gravity of 0.786 at 100°F., and a viscosity of 1.72 centistokes at 100°F.

Temperatures were measured with copper-constantan thermocouples and a multipoint temperature indicator calibrated to ±0.5 F. Readings were taken after the inlet and exit temperatures were stabilized for at least 10 min. at high flow rates and 20 min. at low flow rates. In these studies the difference in the heat balances between the water and kerosene streams varied from 0 to 17%, with the average difference being approximately 3%.

TABLE 2. DISTRIBUTOR DIMENSIONS

Distributor designation	Diameter, in.	No. of holes	Diameter of holes, in.
A	6	280	1/8
B	6	360	1/16
C	6	730	1/32
D	6	330	1/64
E	4	160	1/16

TABLE 4. AVERAGE DROP SIZES

Distributor designation	Average diameter of kerosene droplets, in.
A	0.27
B & *E*	0.22
C	0.13

RESULTS AND DISCUSSION

Heat transfer from kerosene to water was studied in spray towers 4 to 6 in. in diameter and 7 to 10 ft. in height with the use of drop distributors with hole sizes ranging from $\frac{1}{64}$ to $\frac{1}{8}$ in. Regimes of flow included free-rising kerosene droplets, free-falling water droplets, and packed kerosene droplets rising in hindered flow. Experimental data are given in Table 3.*

The average drop sizes for kerosene dispersed in water by the distributors are found in Table 4. These values were obtained with a camera equipped with close-up attachments. The average drop size for each distributor was found not to vary significantly with throughput, ratio of water-to-kerosene flow rates, degree of droplet packing in the column, and column height. Distributor *D* was found to produce immeasurable small drops that emulsified and were easily entrained in the exiting water.

The average drop sizes for water dispersed in kerosene by the distributors were indeterminable because of opacity and rapid coalescence. Distributors *A* and *B* produced large drops that coalesced so rapidly while descending in the column that eventually the water channeled through the kerosene in distinct streams. Distributors *C* and *D* produced drops that would permit the column to operate without channeling, but it was not possible to pack the droplets close together as in hindered settling to attain higher water holdup.

The dispersed phase holdup was obtained by turning off the valves on the inlet streams quickly and permitting the dispersed phase to settle. By controlling carefully the position of the interface and correcting for the excess dispersed phase opposite the distributor, consistent values of holdup were obtained.

Efficiency with Free-Flowing Dispersed Phase

The efficiency of column *A* for transferring heat from free-rising kerosene droplets to water is shown in Figure 3. The column was operated with matched flow-heat capacities

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*Deposited as document 9685 with the American Documentation Institute, Photoduplication Service, Library of Congress, Washington 25, D.C., and may be obtained for $1.25 for photoprints or 35-mm. microfilm.

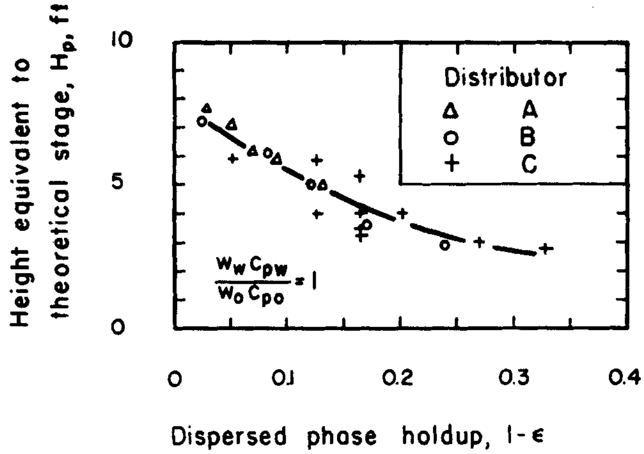

Fig. 3. Effect of holdup on efficiency of column *A* for heat transfer from free-rising kerosene droplets to water.

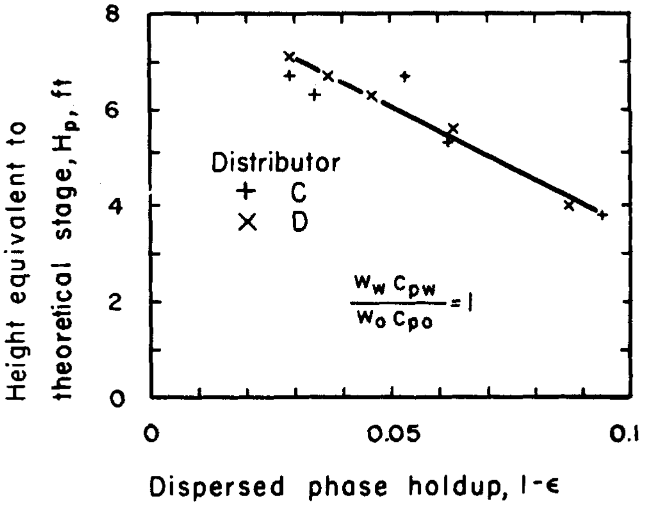

Fig. 5. Effect of holdup on efficiency of column *A* for heat transfer from kerosene to free-falling water droplets.

($w_w c_{pw} = w_o c_{po}$) using distributors *A*, *B*, and *C* to disperse the kerosene. The height per theoretical heat transfer stage, which for this case is equal to the height per heat transfer unit, is seen to decrease with dispersed phase holdup. The column is relatively ineffective for heat transfer in that heights greater than five column diameters are required per theoretical stage.

In addition to analyzing the column performance with the use of the theoretical stage concept, we used the heat transfer coefficients. From Figure 4 the effect of dispersed phase holdup on the overall heat transfer coefficient is noted. As holdup increases, the overall film heat transfer coefficients increase. The distributor producing the largest average drop size yielded the greatest heat transfer coefficients; as the average drop size decreased, the coefficients decreased. This, in part, is due to the increased internal circulation within the larger drops.

It has been proposed by various investigators, such as Sideman and Shabtai (*13*), that the oscillating motion of the drops was the major factor in promoting internal circulation within the drop. It was observed in the present studies that the larger drops tended to be oscillating ellipsoids. As the drops became smaller, they were more spherical and oscillation was almost nonexistent.

The performance of column *A* for transferring heat from kerosene to free-falling water droplets is shown in Figure 5. The column was operated with matched flow-heat capacities ($w_w c_{pw} = w_o c_{po}$) using distributors *C* and *D* to disperse the water. Because of coalescence at higher throughputs, the maximum dispersed phase holdup attainable was about 10%. The height per theoretical stage, correlated as a function of holdup, is seen to be somewhat less than for kerosene dispersed.

Other studies were carried out with free-rising kerosene droplets and free-falling water droplets at varying water to kerosene ratios, that is, unmatched flow-heat capacities.

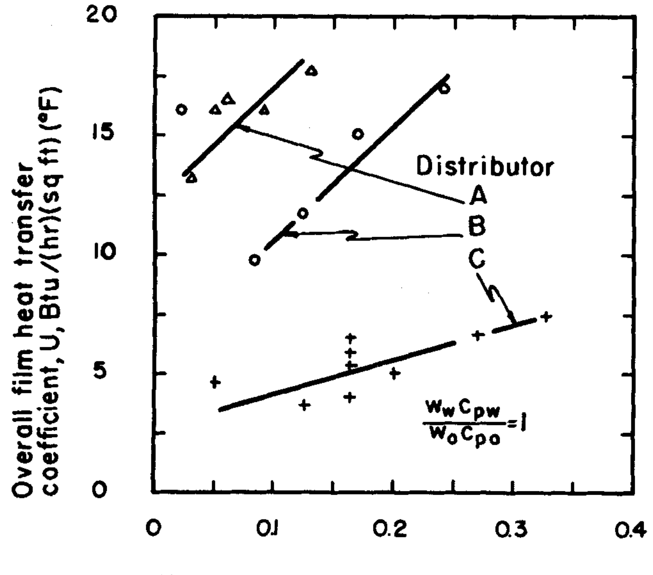

Fig. 4. Effect of holdup on film coefficient for heat transfer from free-rising kerosene droplets to water in column *A*.

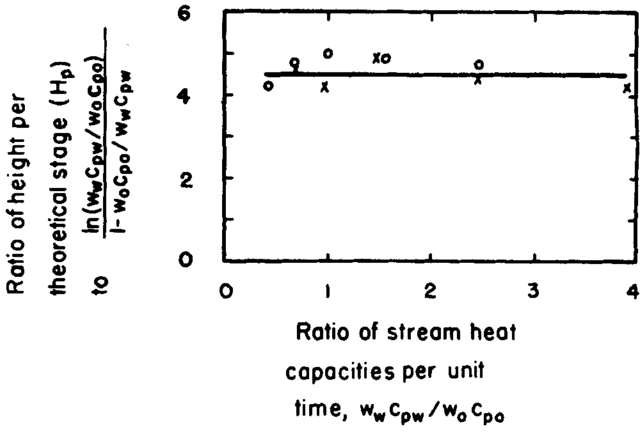

Fig. 6.

	Distributor	Mass velocity of water, lb./(hr)(sq. ft.)	Free-flowing dispersed phase
0	*B*	2,300	Kerosene
X	*D*	3,800	Water

Effect of ratio of flow heat capacities on efficiency of column *A* for heat transfer from kerosene to water.

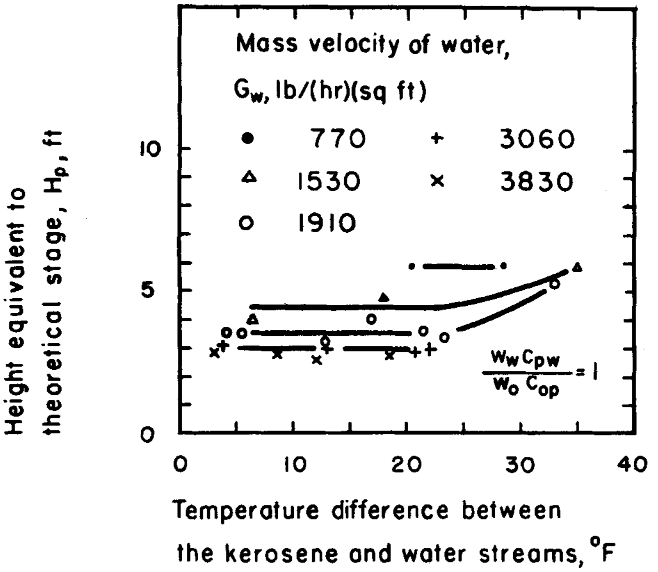

Fig. 7. Effect of thermal driving force on efficiency of column *A* with distributor C for heat transfer from free-rising kerosene droplets to water.

Using as a guide the relationship between the (*UA*) requirement per theoretical heat transfer stage and the ratio of the stream heat capacities per unit time as expressed by Equation (*15*), we developed the correlation shown in Figure 6. The effect on height per theoretical stage [which is inversely proportional to the (*UA*) requirement per theoretical stage for a specific column, drop size, and dispersed phase holdup] of varying driving forces resulting from unbalanced flow-heat capacities is readily accounted for.

A further test of the ability to isolate effects of driving forces from column efficiency is shown in Figure 7. Using column *A* with free-rising kerosene droplets dispersed by distributor *C* such that $w_w c_{pw} = w_o c_{po}$, we see that the height per theoretical heat transfer stage is a function solely of throughput, as indicated earlier in Equation (16), and is independent of ΔT up to values of about 20°F. At higher values of ΔT, the larger axial temperature gradients tend to increase the value of H_p at a given G_w. Since only two points were obtained, additional data are needed to substantiate this result.

Efficiency with Dispersed Phase in Hindered Flow

It was found in the preceding free-rise studies with kerosene distributed that the dispersed phase holdups were limited to approximately 33% by excessive entrainment of kerosene in the exiting water. In order to gain additional increases in column efficiency, it became necessary to increase further dispersed phase holdups. An expanded tapered section of column at the distributor to decrease the velocity of the water entering the downcomers would decrease entrainment and allow the column to be operated at higher throughputs and higher holdups. Another method of increasing dispersed phase holdup is to slow down the coalescing rate of the kerosene droplets and cause the layer of uncoalesced, densely packed kerosene droplets to become larger, as shown in Figure 2, and gradually fill the column.

The latter technique was used in this study with various mats of glass wool as coalescers to attain a range of coalescing rates. In operation, the kerosene feed rate was increased beyond the coalescing rate of the coalescer in use in order to build up the layer of compacted kerosene droplets. When the column was filled with packed kerosene droplets, the kerosene rate was decreased to the normal coalescing rate and the interface between the free-rise zone and the packed zone was maintained immediately above the distributor with only small fluctuations. It is expected that an expanded column diameter at the distributor would have made control of the position of the free-to-packed interface less dependent on flow rate.

The height per theoretical heat transfer stage for columns *A*, *B*, and *C*, using distributors with $\frac{1}{16}$-in. holes to

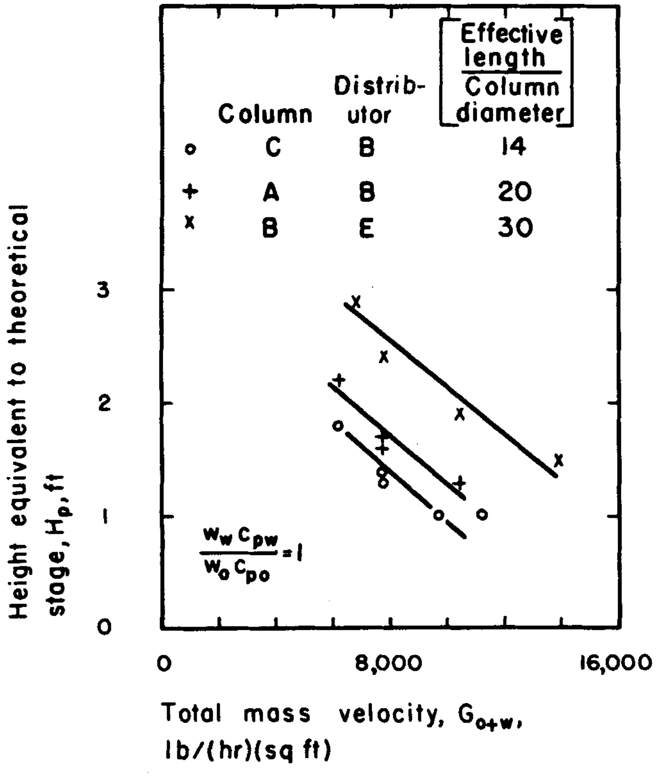

Fig. 8. Effect of thoughput on column efficiency for heat transfer from packed kerosene droplets to water.

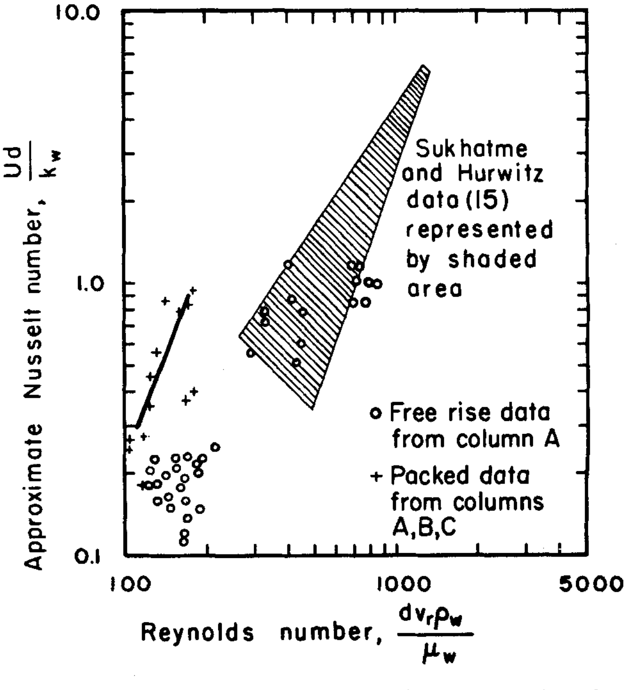

Fig. 9. Correlation of film coefficient for heat transfer from free-rising and packed kerosene droplets to water.

disperse the kerosene and coalescers to hinder the flow of the kerosene droplets to pack them, is shown in Figure 8. Values of H_p ranging from 1 to 3 ft. under packed conditions with dispersed phase holdups of 52 to 84% are greatly improved over the 3 to 8 ft., attainable under free-rise conditions with holdups of 2 to 33%.

Attempts to increase the efficiency of heat exchangers in which water is dispersed in kerosene by packing closer together the droplets of water were unsuccessful because of coalescence.

Scale-Up

The data in Figure 8 show that the column length and diameter affect the height equivalent to a theoretical stage for heat transfer from densely packed kerosene droplets to water. It would therefore be advisable to use some other rate expression for scaling up spray heat exchangers.

Various investigators have concluded that the Nusselt number is a function of the Reynolds number, often in combination with the Prandtl number. Exact relationships between the Nusselt and Reynolds numbers are well summarized by Sideman (*14*). Hence a generalized correlation based on the overall film coefficient is shown in Figure 9 relating an approximate Nusselt number to a Reynolds number. Figure 9 pertains to heat transfer from free-rising kerosene droplets and heat transfer from packed kerosene droplets in hindered flow to water. The values of overall film coefficient are comparable in both types of flow, and range from 4 to 23 B.t.u./(hr.)(sq. ft.)(°F.).

The properties of water were used in the correlation; however it was not certain in all cases whether it was the controlling resistance. From similar studies carried out by Sukhatme and Hurwitz (*15*), it can be shown in this work that for the case of kerosene drops approximately 0.13 in. in diameter (distributor *C*), water is the controlling phase. For larger kerosene drops no such conclusion can be reached.

The free-rise data in Figure 9 fell within and extended the data of Sukhatme and Hurwitz (*15*). The Nusselt numbers in the packed studies were comparable to those in the free-rise studies but corresponded to lower Reynolds numbers.

CONCLUSIONS

For heat transfer from kerosene to water in liquid-liquid spray towers, dispersed phase holdup has the greatest effect on column efficiency. As the holdup of dispersed kerosene is increased from 2 to 84%, the efficiency of heat transfer columns is increased by a factor of five or more. The drop holdup attainable with water dispersed was limited to 10% because of coalescence, and thus such heat exchangers have less potential than those with kerosene dispersed.

It was possible to correlate data from columns of varying diameter and height for purposes of scale-up, but data scatter becomes greater as correlations become more generalized.

ACKNOWLEDGMENT

The authors appreciate the interest and advice of Professor Merrell R. Fenske, and are grateful for the support of the Mobil Foundation Inc., and the E. I. du Pont de Nemours and Company, Inc.

NOTATION

a = interfacial heat transfer area per unit volume, sq. ft./cu. ft.
A = interfacial heat transfer area, sq. ft.
c_p = heat capacity, B.t.u./(lb.)(°F.)
d = average diameter of dispersed phase droplets, ft.
h = individual film heat transfer coefficient, B.t.u./(hr.)(sq. ft.)(°F.)
G = mass velocity, lb./(hr.)(sq. ft.)
H = enthalpy, B.t.u./lb.
H_p = height equivalent to theoretical heat transfer stage, ft.
k = thermal conductivity, B.t.u./(hr.)(sq. ft.)(°F./ft.)
n = number of drops of dispersed phase in column
N_{Nu} = Nusselt number
N_p = number of theoretical heat transfer stages
N_{Re} = Reynolds number
N_{t0} = number of overall heat transfer units
q = heat transfer rate, B.t.u./hr.
T = temperature, °F.
ΔT = temperature difference between the fluids, °F.
U = overall heat transfer coefficient, B.t.u./(hr.)(sq. ft.)(°F.)
(Ua) = volumetric heat transfer coefficient, B.t.u./(hr.)(cu. ft.)(°F.)
v_r = relative velocity between the fluids, ft./sec.
v_s = superficial velocity, ft./sec.
V = volume of column effective in heat transfer, cu. ft.
w = mass rate of flow, lb./hr.

Greek Letters

ε = holdup of continuous phase; fraction of effective tower volume occupied by continuous phase
μ = viscosity, lb./(sec.)(ft.)
ρ = density, lb./cu. ft.

Subscripts

a = bottom of heat exchanger
b = top of heat exchanger
c = controlling phase
m = mean; average
o = oil phase
w = water phase

Superscript

$*$ = equilibrium value

LITERATURE CITED

1. Bauerle, G. L., and R. C. Ahlert, *Ind. Eng. Chem. Process Design Develop.*, 4, No. 2, 225-31 (1965).
2. Bray, U. B., C. E. Swift, and D. E. Carr, *Proc. Am. Petrol. Inst.*, Sec. 3, 96 (1933).
3. Elzinga, E. R., Jr., and J. T. Banchero, *Chem. Eng. Progr. Symp. Ser. No. 29*, 55, 149-161 (1959).
4. "Feasibility, Research, and Analysis of Heat Exchange Systems Without Metallic Heat Exchange Surfaces and Laboratory Study of Single Stage Vapor Reheat Distillation," PB181299, Office Saline Water, U. S. Dept. Interior (1962).
5. Garwin, L., and B. D. Smith, *Chem. Eng. Progr.*, 49, No. 11, 591-602 (1963).
6. Geoffrin, P. L., M. S. thesis, Pennsylvania State Univ., University Park (1964).
7. Grover, S. S., and J. G. Knudsen, *Chem. Eng. Progr. Symp. Ser. No. 17*, 51, 7108 (1955).
8. Hartland, S., *Trans. Inst. Chem. Engrs.*, 44, T116-121 (1966).
9. Johnson, A. I., G. W. Minard, C. Huang, J. H. Hansuld, and V. M. McNamara, *AIChE J.*, 3, No. 1, 101-110 (1957).
10. Letan, Ruth, and Efraim Kehat, *ibid.*, 5, 804 (1965).
11. Pierce, R. D., O. E. Dwyer, and J. J. Martin, *ibid.*, 5, No. 2, 257-62 (1959).
12. Sagar, D. V., T. K. Ramanujam, V. Krishnamurty, and C. Venkata-Rao, *Indian J. Technol.*, 3, 79-82 (1965).
13. Sideman, Samuel, and H. Shabtai, *Can. J. Chem. Eng.*, 42, No. 3, 107-117 (1964).
14. Sideman, Samuel, "Advances in Chemical Engineering," Vol. 6, Chap. 4, Academic Press, New York (1966).
15. Sukhatme, S. P., and M. Hurwitz, "Heat Transfer in the Liquid-Liquid Spray Tower," PB166315, Office Saline Water, U. S. Dept. Interior (1964).
16. Wilke, C. R., C. T. Cheng, V. L. Ledesma, and J. W. Porter, *Chem. Eng. Progr.*, 59, No. 12, 69-75 (1963).
17. Woodward, T., *Chem. Eng. Progr.*, 57,No. 1, 52-57 (1961).

Manuscript received December 28, 1966; revision received April 4, 1967; paper accepted April 6, 1967.