

HEAT TRANSFER IN AN OIL-STEAM DIRECT-CONTACT
CONDENSATION SYSTEM

A Thesis

by

CHRISTOPHER ALEXANDER MAIN

Submitted to the Graduate College of
Texas A&M University in
partial fulfillment of the requirement for the degree of
MASTER OF SCIENCE

August 1979

Major Subject: Chemical Engineering

HEAT TRANSFER IN AN OIL-STEAM DIRECT-CONTACT
CONDENSATION SYSTEM

A Thesis

by

CHRISTOPHER ALEXANDER WAIN

Approved as to style and content by:

R. P. Davison

(Chairman of Committee)

C. Holland

(Head of Department)

Melden J. Fudge

(Member)

R. S. Anthony

(Member)

August 1973

437036

ABSTRACT

Heat Transfer in an Oil-Steam Direct-Contact
Condensation System. (August 1973)

Christopher Alexander Waln, B.S., Texas A&M University

Directed by: Dr. R. R. Davison

The volumetric heat transfer coefficients associated with direct-contact oil-steam condensation were obtained by calculating enthalpy balances on a three stage packed condensation system. The system was run at a variety of physical conditions and with two different packing types.

The volumetric heat transfer coefficients were correlated with packing factor, oil and water flow, steam density, and liquid to vapor ratio. This correlation was obtained by the least squares method. It was additionally determined that a chemical modification of the oil used improved heat transfer at low water to oil ratios.

The results of this investigation tend to indicate that direct-contact oil-steam condensation could be competitive with more conventional methods of steam condensation.

ACKNOWLEDGMENTS

This investigation was made possible through the suggestion and direction of Dr. R. R. Davison.

To Dr. R. G. Anthony and Dr. M. J. "Bob" Fox, Jr., I wish to extend grateful acknowledgment for their contributions to my research efforts and thesis.

For their support by financial aid, the Texas Engineering Experiment Station and Dr. C. D. Holland are acknowledged.

Finally, this thesis is dedicated to all those men who are still listed as Missing In Action in Southeast Asia.

TABLE OF CONTENTS

	PAGE
ABSTRACT.....	iii
ACKNOWLEDGMENTS.....	iv
LIST OF TABLES.....	vi
LIST OF FIGURES.....	vii
CHAPTER	
I. INTRODUCTION.....	1
II. PREVIOUS WORK.....	3
III. EXPERIMENTAL SYSTEM AND ITS OPERATION..	7
Experimental System.....	7
Operation.....	16
IV. DERIVATIONS.....	21
V. RESULTS AND DISCUSSION.....	32
NOMENCLATURE.....	45
BIBLIOGRAPHY.....	47
APPENDIX	
I. CONVERSION FACTORS.....	48
II. DATA USED IN HEAT LOSS CALCULATIONS....	49
III. DATA CORRELATION AND STATISTICS.....	52
VITA.....	56

LIST OF TABLES

TABLE		PAGE
I.	DATA: UNMODIFIED OIL-PALL RINGS.....	33
II.	DATA: UNMODIFIED OIL-INTALOX [®] SADDLES.....	36
III.	DATA: AVERAGE MODIFIED OIL-INTALOX [®] SADDLES.....	38
IV.	CONVERSION FACTORS.....	48
V.	TEMPERATURE DIFFERENTIALS USED IN HEAT LOSS CALCULATIONS.....	51
VI.	EXPERIMENTAL AND CORRELATED VOLUMETRIC HEAT TRANSFER COEFFICIENTS.....	53
VII.	CURVE-FIT STATISTICS.....	55

LIST OF FIGURES

FIGURE		PAGE
1.	DIRECT-CONTACT CONDENSING SYSTEM.....	8
2.	TYPICAL DESUPERHEATER.....	9
3.	TYPICAL CONDENSING STAGE.....	12
4.	SEPARATOR SYSTEM.....	14
5.	MASS FLOW.....	22
6.	ENTHALPY FLOW.....	23
7.	ENTHALPY FLOW FOR $Q_{Loss,n}$ DETERMINATION.....	27
8.	TEMPERATURE APPROACH PROFILE FOR STAGE n.....	28
9.	TEMPERATURE APPROACH PROFILE (HARRIOTT AND WIRGANDT 1964).....	29
10.	BEST VOLUMETRIC HEAT TRANSFER COEFFICIENTS.....	40

CHAPTER I

INTRODUCTION

Today most commercial desalination is done by flash distillation. This process has three drawbacks. Heat transfer across the metal surfaces employed is limited due to limited area. The heat transfer may be further impeded by the build-up of insoluble scale on the transfer surface. This scaling is also responsible, along with corrosion, for inherently high maintenance costs. Lastly, the necessity of corrosion resistant construction materials causes initial equipment costs to be high.

Direct-contact flash distillation is another method of seawater desalting capable of meeting the high freshwater production levels usually associated with conventional flash distillation. The direct-contact process avoids the drawbacks of the conventional process. The major disadvantage in the direct-contact process is the necessity of phase separation. In keeping with the cost advantages of the direct-contact flashing of briny and brackish feedwaters, direct-contact heat transfer could

The citations on the following pages follow the style of the AIChE Journal.

be used to advantage in the condensation of steam from the direct-contact flash.

Direct-contact condensation has received little attention in the literature. When direct-contact condensation of steam in an industrially feasible system is considered, this is especially true. Since the available literature were both limited and subject to contradiction, a broadly based program of study was planned. This program included the design, construction, and operation of a pilot-plant scale direct-contact oil-steam condensation system. The construction materials used were not what would be used in a full scale system. However, the heat transfer liquid and physical operating conditions were those which would be used on an industrial scale.

An effort was made to obtain a comprehensive data field by operating the system over a wider range of physical conditions. Two different packing types were used. When the second set of packing materials was used, the heat transfer liquid was used both in its pure form and modified by the addition of an aliphatic alcohol.

The final objective of this work was the correlation of the collected data in such a fashion that larger scale systems could be designed.

CHAPTER II

PREVIOUS WORK

As mentioned before, the available literature on direct-contact condensation is both limited and subject to contradiction. This is particularly true of work done on packed condensers operated at industrially feasible conditions. This is not to say, however, that what information is available is of diminished value.

Cheng (1963) and Harriott and Wiegandt (1964) all worked on direct-contact heat transfer in packed condenser systems. The volumetric heat transfer coefficients obtained by Cheng (1963) were, however, substantially less than those developed by Harriott and Wiegandt (1964). Working in conjunction with Wilke *et al.* (1964), Cheng (1963) used Aroclor, a highly viscous, expensive, chlorinated bi-phenyl as a heat transfer liquid. This was done because Wilke's group was using Aroclor as a heat transfer liquid in the flash distillation of seawater, and it was desired to combine the two processes for the purposes of cost reduction and heat economy.

Cheng (1963) found that the heat transfer coefficients, characterized by HTU, obtained in his

work could be reasonably predicted by a modification of the Chilton-Colburn J factor method:

$$(\text{HTU})_L/d_p = C(\text{Re.})_L^d (\text{Pr.})_L^f (\text{Ga})_L^h \quad (1)$$

These volumetric heat transfer coefficients, of approximately $111\,800 \text{ J/m}^3 \cdot \text{s} \cdot \text{K}$, were well below, by a factor of four, the values predicted by the penetration theory supported by the work of Harriott and Wiegandt (1964). They have suggested that, possibly, the very high viscosity of Aroclor was responsible for the reduced values obtained by Cheng (1963). Their own work with a water and methyl-chloride system tends to support this claim.

Allowing for this variation, however, both groups agreed that heat transfer in packed condensers is controlled by the liquid phase resistance. Harriott and Wiegandt (1964) also observed that the volumetric heat transfer coefficient varied with liquid flow, reporting the relationship:

$$U_v \propto L^{0.4 \text{ to } 0.6} \quad (2)$$

It was also the general consensus that the volumetric heat transfer was independent of gas flow. Harriott and Wiegandt (1964) reported heat transfer

coefficients of $2\,795\,000\text{ J/m}^3\cdot\text{s}\cdot\text{K}$ on their system well before flooding conditions were reached. Whereas Cheng (1963) only achieved $111\,800\text{ J/m}^3\cdot\text{s}\cdot\text{K}$ rather than the $465\,800\text{ J/m}^3\cdot\text{s}\cdot\text{K}$ predicted for his system by the penetration theory (Harriott and Wiegandt 1964).

Cheng's work with Wilke *et al.* (1964) was limited by the use of Aroclor which is industrially unfeasible due to its high initial purchase cost, pumping costs, limited heat transfer ability, and odor and flavor which tend to taint the product freshwater.

Lackey (1961) also used Aroclor. However, he used a venturi contractor rather than a packed column to condense steam. In order to produce volumetric heat transfer coefficients of $2\,795\,000 - 7\,450\,000\text{ J/m}^3\cdot\text{s}\cdot\text{K}$, he found it necessary to use temperature gradients of 5.55, 22.2, and 38.9 degrees kelvin. While these values are obviously impractical on an industrial scale, their use did emphasize the dependence of steam flow rate on temperature differential. To obtain an increase in the steam flow rate without using these excessive temperature gradients, Hsu and Hickman (1972) have suggested the addition of "sensitizers" to the heat transfer liquid. Added in amounts less than one per cent by weight, these "sensitizers", notably aliphatic alcohols,

provide surface sites for condensation. However, these alcohols will not increase the heat transfer to levels beyond that which the heat transfer liquid is inherently capable.

Cheng (1963), Harriott and Wiegandt (1964), Sideman (1966), and Maa and Hickman (1972) all agree on the basic characteristics of direct-contact heat transfer in packed condensers. The results of their work, however, leave considerable doubt as to how direct-contact condensation compares with other methods of condensation. Sideman (1966), in his compilation of work directed at this subject, stated, "...it is noted that, despite the advantages and effectiveness realized in utilizing latent heat transport, relatively little is known of the basic mechanism of this important phenomenon, and the various factors affecting it." Seven years later this is still the case.

CHAPTER III

EXPERIMENTAL SYSTEM AND ITS OPERATION

Experimental System

The system shown in Figure 1 (p. 8) is a direct-contact oil-steam condensation system. It consists of three sub-systems. One is for steam supply, one for condensation, and one for phase separation. In addition, facilities are provided such that any three consecutive stages of a thirty to forty stage system may be modeled.

The heat transfer liquid used is a hydrocarbon fraction, approximately ninety per cent of which boils within 2 degrees kelvin of the C_{13} boiling point. This oil was selected for its low viscosity, low expense, water insolubility, low toxicity level, and its inability to act as a solvent for the salts associated with seawater.

The apparatus shown in Figure 2 (p. 9) is a typical desuperheater in the steam supply sub-system. The steam supply sub-system consists of three desuperheaters and their associated control elements. Each of the desuperheaters is made of a section of six-inch

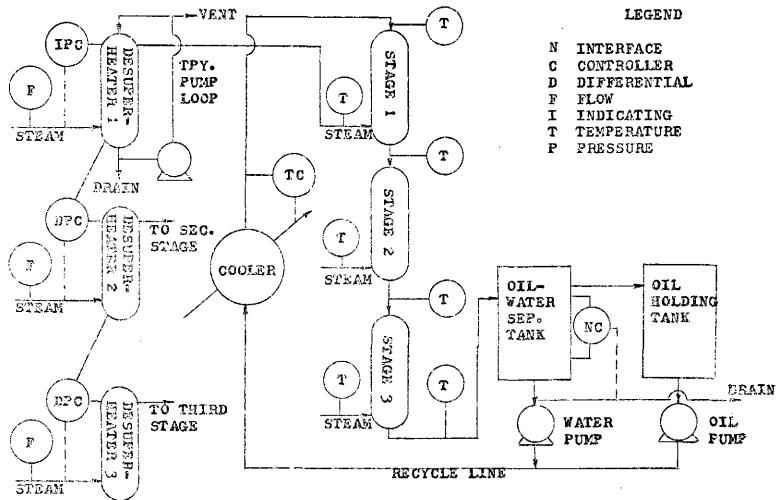


FIGURE 1. DIRECT-CONTACT CONDENSING SYSTEM

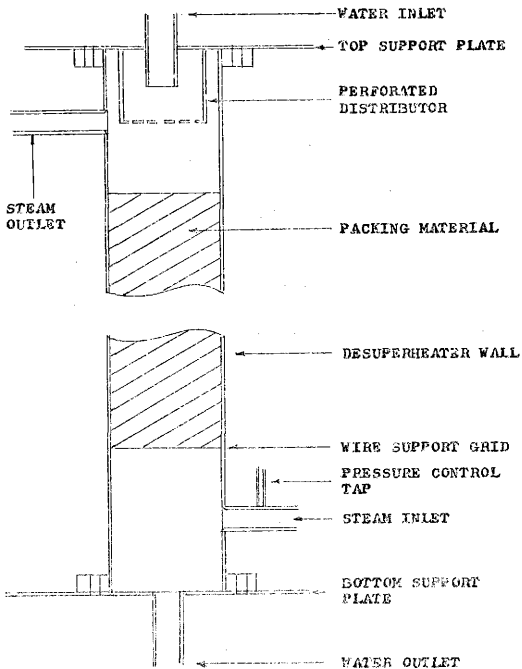


FIGURE 2. TYPICAL DESUPERHEATER

steel pipe 0.915 m long. A one-fourth of an inch wire mesh grid, 0.152 m above the lower end of the desuperheater, supports 0.610 m of wet-packed one-half inch Raschig rings. A hot water circulating pump recycles water at a high flow rate around a one-inch pipe loop from the bottom of the desuperheater to the perforated steel distributor above the packing material. Utility steam at a pressure of 689 475 N/m² guage is supplied to each desuperheater through a pneumatic control valve. The steam enters the desuperheaters from the side, 0.076 m below the packing support grid. Each pump loop is supplied with a manual drain and a pressure relief valve. The desuperheaters are each insulated with one-inch split-shell glass fiber pipe insulation.

Desuperheater 1 is on absolute pressure control. Separate Honeywell pressure transmitters are used for vacuum operation and operation at atmospheric pressure and above. The transmitters are connected to a Honeywell differential pressure transmitter, which is connected to a Honeywell three pen recorder for flow measurement. Desuperheaters 2 and 3 are essentially identical in their control arrangement. Instead of direct pressure control, however, desuperheaters 2 and 3 are on differential pressure control with the desuperheater immedi-

ately above; see Figure 1 (p. 8).

Figure 3 (p. 12) characterizes a typical packed stage in the condensing sub-system. Each of these sections is connected to the corresponding desuperheater by a one-inch steel pipe with a manual shut-off valve and a one-way check valve. The condensation sections consist of a steel distributor section, a glass pipe packed section, and a glass pipe support and steam supply section. These sections are insulated in the same manner as the desuperheaters.

The distributor section has four one-inch pipe downcomers with V-notched tops to provide for even flow distribution. In stages 2 and 3, liquid from the stage above comes into the distributor section via a 1.22 m long section of four-inch steel pipe which acts as a hydraulic leg. In the first stage, liquid enters the distributor section from the phase separator recycle line. Each distributor section is designed to prevent vapor flow between stages. The distributor sections are fabricated from a 0.152 m long section of six-inch steel pipe. Each of these sections is provided with an air bleed valve.

A typical glass pipe packed section is 0.61 m long and 0.152 m in diameter. It is completely packed with

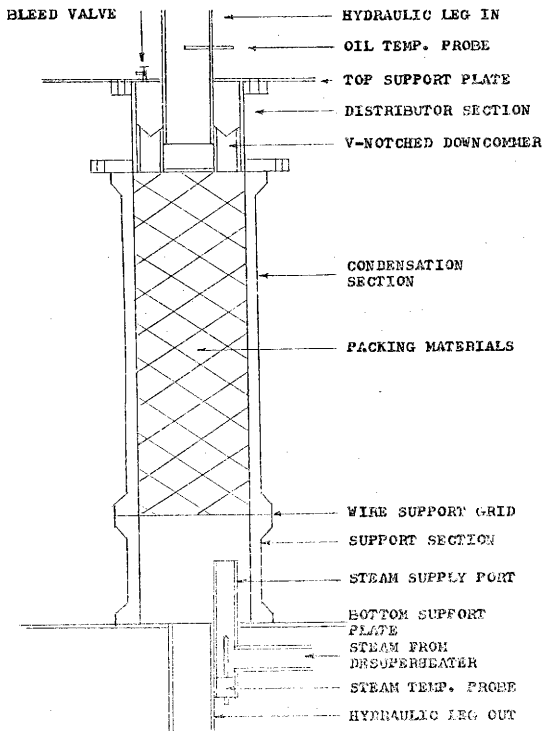


FIGURE 3. TYPICAL CONDENSING STAGE

either five-eighths of an inch steel Fall rings or three-fourths of an inch ceramic Intalox H saddles.

The glass pipe support sections include a one-fourth of an inch wire mesh grid at the top end for the support of the packing materials. The steam supply lines from the desuperheaters enter through the support section end plates. Each support section is constructed from a 0.152 m long section of six-inch glass pipe.

The liquid flow from the last condensation stage enters the phase separation sub-system by four-inch steel pipe. The phase separation sub-system in Figure 4 (p. 14) consists of two insulated 0.189 m^3 steel tanks. These tanks are arranged in series and are connected by a one-inch steel pipe oil overflow line. The first tank acts as a gravity separator. A Honeywell differential pressure transmitter connected to an indicating controller and pneumatic valve provides oil-water interface control in the separator tank. Accumulated water is discharged from the tank by the interface control valve. A water recycle line runs from the lower end of the separator tank through a centrifugal pump and back to the condensation sub-system. In the recycle line there is an integral orifice connected to a Honeywell recorder-controller for water flow control.

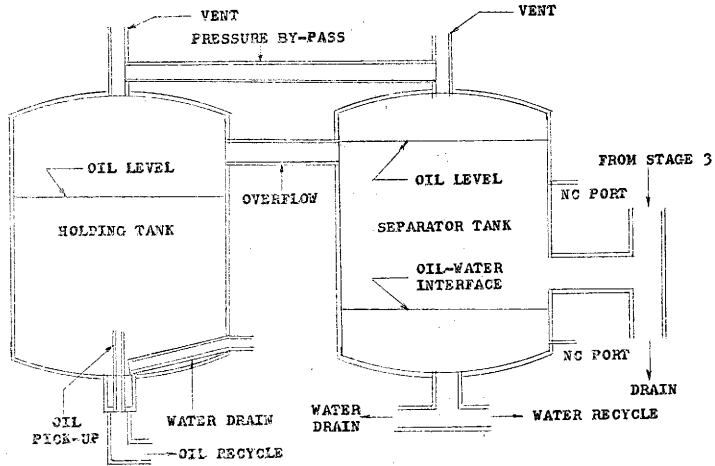


FIGURE 4. SEPARATOR SYSTEM

From the separator tank, oil overflows into the second tank, which is provided to give sufficient oil hold-up to prevent oil recycle from getting ahead of phase separation. This tank is provided with a manual system to remove water accidentally carried over from the separator tank and thus to prevent water from entering the oil recycle line. Oil is recycled by means of a Flowthru [®] pump. Downstream from the pump is the same type flow-measuring system that is used on the steam lines. The exception in this case is that a flow recorder-controller is used to control the oil flow rate. The oil and water recycle lines merge prior to entering an over-sized, stainless steel, shell and tube heat exchanger. The recycle stream flows tube side, while 551 580 N/m² guage utility water flows shell side. This heat exchanger provides temperature control for the recycle stream. The temperature sensor is connected to a Honeywell temperature recorder-controller and is located in the inlet to the distributor section for condensation stage 1. Yellowstone Industries thermistors are located as shown in Figure 1 (p. 8). These thermistors provide signals which enter a resistance balance. This balance is used in conjunction with calibration tables to obtain temperature readings.

Operation

Data runs with the system described were made at pressures corresponding to those of saturated steam at the temperatures of approximately 339 K, 356 K, 374 K, and 389 K. At each of these temperatures, a series of oil flow rates was run to vary the loading of the column. At each oil flow rate, a series of water flow rates was run to model various groups of three stages throughout a thirty to forty stage system.

This general format was followed with the combinations of Pall rings and pure oil, Intalox [®] saddles and pure oil, Intalox [®] saddles and oil modified with 0.5 per cent octadecanol, and Intalox [®] saddles and oil modified with 1.0 per cent octadecanol. This scheme of operation provided a wide data field.

Some variations were required when operating under vacuum; however, the operating procedure used at ambient pressure and above is fairly indicative of how the data was obtained. The following procedure was used:

1. The oil recycle pump was operated at full flow for five minutes to establish vapor seals in the condensation stage distributors and hydraulic legs.

2. The main steam supply valve was opened so that condensate could be purged through the steam trap.
3. The indicating controller for desuperheater 1 was set at the desired value. The manual supply valve to the desuperheater was then opened. The pump in the circulating loop was turned on.
4. When the pressure in desuperheater 1 reached the desired value, the pressure differential to desuperheater 2 was set. This differential was not allowed to exceed the pressure exerted by 1.22 m of oil in the hydraulic leg connecting condensing sections 1 and 2. The manual supply valve to desuperheater 2 was then opened with the circulating pump turned on.
5. When the desired pressure was reached in desuperheater 2, step four was repeated for the last desuperheater.
6. When all three desuperheaters were operating in a stable manner, the air bleed valve for condensing section 1 and the supply valve from desuperheater 1 were opened.
7. As soon as steam began to vent from the bleed valve, it was shut, and the procedure in step

- six was repeated for condensing sections 2 and 3.
8. The separator interface control was set for an interface 0.25 m above the tank bottom.
 9. The oil and water flow controllers were set at zero flow. The temperature controller was set at six degrees kelvin below the desired value.
 10. The oil flow pump was then turned on. Oil flow was increased at the rate of one controller division per minute until the desired value was reached.
 11. Five minutes after reaching the desired oil flow rate, the water flow pump, if it was to be used, was turned on. The water flow rate was then adjusted upward at the rate of one controller division per minute until the desired value was reached.
 12. Five minutes after the correct water flow rate was established, the temperature controller setting was increased at the rate of one degree kelvin per minute until the desired reading was reached.
 13. When all recorders indicated steady-state operation had been obtained, the thermistor

resistance readings were measured. When three consecutive sets of temperature readings were within 0.05 degree kelvin, the last set of temperatures was recorded.

14. Changes in the oil and water flow rates for further data collection were accomplished at the same rates that the initial settings were made.
15. Changes in the operating pressures were guided by intuition to prevent system upset.

Shutdown procedures were simple. The following was the procedure used:

1. Steam flow was stopped to each desuperheater by setting its pressure controller to zero scale. The manual valves in the supply lines were then closed. The bleed valves for the desuperheaters were then opened with the circulating pump shut off.
2. Condensing section steam supply lines were closed. Oil and water flow controllers were set at the ambient temperature. When temperatures reached 310 degrees kelvin, flow controllers were set at zero flow, and the recycle pumps were shut off.

3. Condensation section bleed valves and phase separator bleed valves were opened, and the main steam valve was shut off.

As water accumulated in the desuperheaters it was necessary to remove it so that the desuperheater packing would not flood. It was also necessary to remove excess condensate from the separator tank and the oil holding tank. At ambient pressures and above, this was simply a matter of opening a valve; however, under vacuum a different method was required. A tank was provided at higher vacuum than that of the system. This tank was connected to the various drain lines as required. When it became necessary to drain one of the various sub-system elements, the drain valve was opened and water drained into the high vacuum tank. The tank was then vented, drained, and pumped back down to the required vacuum.

Prior to data collection runs, hot water, at temperatures corresponding to those used in data collection runs, was run through the system with the steam off. The resulting temperature differences were used to determine heat losses from the system. These heat losses were used in enthalpy balances during data reduction.

CHAPTER IV

DERIVATIONS

In order to obtain volumetric heat transfer coefficients from the flow and temperature data collected, it was necessary to derive mass and enthalpy balances on the system. These balances were then coded in Fortran IV, and the associated calculations were performed by a Digital Scientific META IV computer.

For simplicity in developing the mass and enthalpy balances, a single condensation stage, designated as stage n and bounded by condensation stages $n-1$ and $n+1$, was specified. Figure 5 (p. 22) shows the mass flow scheme for condensation stage n . The mass balance for stage n was

$$m_{o,n-1} + m_{w,n-1} + m_{s,n} = m_{o,n} + m_{w,n} \quad (3)$$

Since no change in the oil mass flow occurred from one stage to the next, equation 3 was further simplified to

$$m_o + m_{w,n-1} + m_{s,n} = m_o + m_{w,n} \quad (4)$$

In the development of the enthalpy balance for stage n , the enthalpy flow scheme shown in Figure 6 (p. 23) was

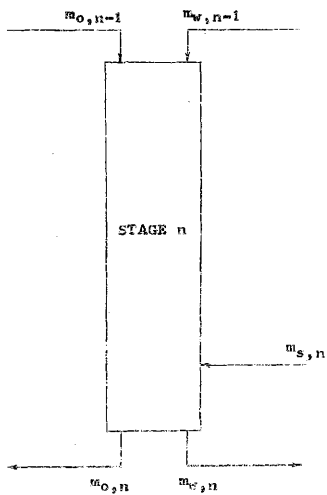


FIGURE 5. MASS FLOW

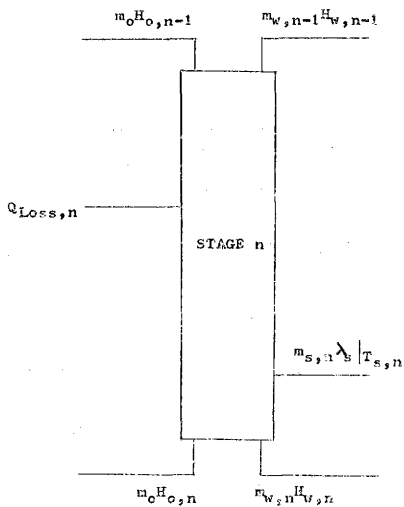


FIGURE 6. ENTHALPY FLOW

used. The assumption was made that the oil and water entering stage n were at the same temperature, T_{n-1} , and that the oil and water leaving stage n were at the same temperature, T_n . It was further assumed that the saturated steam leaving desuperheater n did not change in quality prior to entering condensation stage n . Thus, the heat content of the steam entering the condensation stage was assumed to be the latent heat of saturated steam at the desuperheater temperature and pressure. The mean heat capacity of water was evaluated at the appropriate temperatures by the curve-fit published in the 1957 ASME Steam Tables. This fit has a base temperature of 273 degrees kelvin. The oil mean heat capacity was evaluated at the proper temperatures by the estimation method suggested by Cragoe in 1929. This fit also has a base temperature of 273 degrees kelvin and is shown below:

$$\bar{c}_p = \frac{0.4924 + 0.00081 T^{\circ}\text{C}}{(560)^{0.5}} \quad (5)$$

Based on these assumptions, the enthalpy balance for condensation stage n was developed.

With the further assumption that the system was in steady-state operation, the following is true:

$$\begin{aligned}
 m_o H_{o,n-1} + m_{w,n-1} H_{w,n-1} + m_{s,n} \lambda_s |_{T_{s,n}} &= \\
 m_o H_{o,n} + m_{w,n} H_{w,n} + Q_{Loss,n} & \quad (6)
 \end{aligned}$$

This equation may be restated for purposes of calculation as

$$\begin{aligned}
 m_o \bar{C}_{Po} |_{T_{n-1}} \cdot T_{n-1} + m_{w,n-1} \bar{C}_{Pw} |_{T_{n-1}} \cdot T_{n-1} + \\
 m_{s,n} \lambda_s |_{T_{s,n}} = m_o \bar{C}_{Po} |_{T_n} \cdot T_n + m_{w,n} \bar{C}_{Pw} |_{T_n} \cdot T_n + \\
 Q_{Loss,n} & \quad (7)
 \end{aligned}$$

Assuming the heat loss term $Q_{Loss,n}$ could be estimated, equation 7 was rearranged in the following manner, so that the steam enthalpy, $Q_{steam,n}$ could be used as a convenient check on steady-state operation:

$$\begin{aligned}
 m_o \bar{C}_{Po} |_{T_{n-1}} \cdot T_{n-1} + m_{w,n-1} \bar{C}_{Pw} |_{T_{n-1}} \cdot T_{n-1} - \\
 (m_o \bar{C}_{Po} |_{T_n} \cdot T_n + m_{w,n} \bar{C}_{Pw} |_{T_n} \cdot T_n + Q_{Loss,n}) = \\
 m_{s,n} \lambda_s |_{T_{s,n}} & \quad (8)
 \end{aligned}$$

Collecting terms in equation 8 gave the following:

$$\begin{aligned}
 m_o \cdot (\bar{C}_{Po} |_{T_{n-1}} \cdot T_{n-1} - \bar{C}_{Po} |_{T_n} \cdot T_n) + \\
 m_{w,n-1} \bar{C}_{Pw} |_{T_{n-1}} \cdot T_{n-1} - m_{w,n} \bar{C}_{Pw} |_{T_n} \cdot T_n - Q_{Loss,n} = \\
 Q_{steam,n} & \quad (9)
 \end{aligned}$$

Prior to the calculation of volumetric heat transfer coefficients, it was necessary to estimate values for $Q_{Loss,n}$. Figure 7 (p. 27) shows the enthalpy flow scheme used in the determination of $Q_{Loss,n}$. It is essentially the same as that used in a standard data run, with the exception that steam and oil enthalpies were not involved. Once again, assuming steady-state operation, the following was derived:

$$\begin{aligned} m_{w,n-1} \bar{C}_{Pw} |_{T_{n-1}} \cdot T_{n-1} &= \\ m_{w,n} \bar{C}_{Pw} |_{T_n} \cdot T_n - Q_{Loss,n} & \quad (10) \end{aligned}$$

However, since the water mass flow is constant from stage to stage, equation 10 may be reduced to

$$m_w (\bar{C}_{Pw} |_{T_{n-1}} \cdot T_{n-1} - \bar{C}_{Pw} |_{T_n} \cdot T_n) = Q_{Loss,n} \quad (11)$$

Once heat loss data was reduced by equation 11, a curve-fit for $Q_{Loss,n}$ was developed. This curve-fit was subsequently used in the calculation of the volumetric heat transfer coefficients.

The final assumption required to calculate the volumetric heat transfer coefficients was the temperature approach profile. Figure 8 (p. 28) shows the profile that was used. Harriott and Wiegandt (1964) indicated the profile shown in Figure 9 (p. 29) for

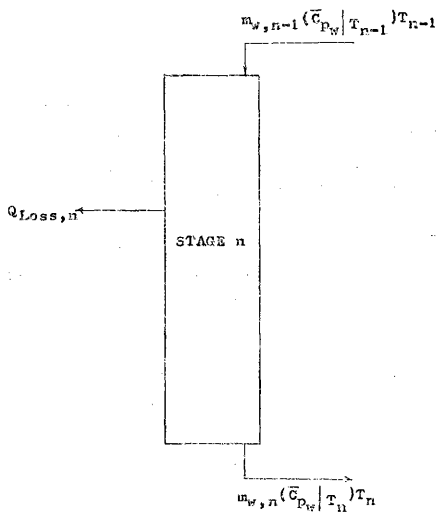


FIGURE 7. ENTHALPY FLOW FOR $Q_{Loss,n}$
DETERMINATION

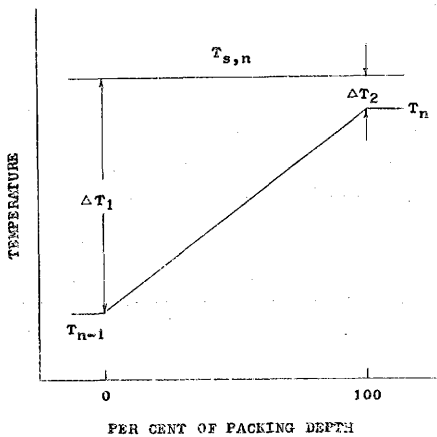


FIGURE 8. TEMPERATURE APPROACH PROFILE
FOR STAGE n

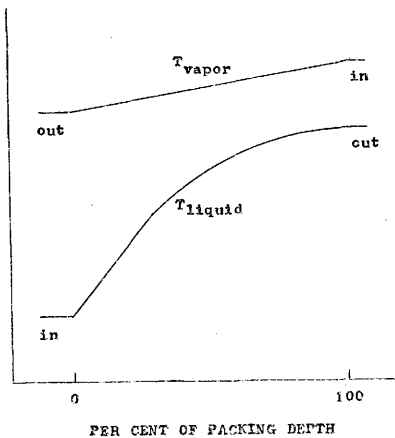


FIGURE 9. TEMPERATURE APPROACH PROFILE
(HARRIOTT AND WIEGANDT 1964)

packed condensation columns. The slight reduction in vapor temperature was due to packing pressure drop. However, without facilities to measure packing internals, the profile chosen was a good approximation. The mean temperature used with this profile was given in Kreith (1966), on page 492, as

$$\Delta T_{1m} = (\Delta T_2 - \Delta T_1) / (\ln(\Delta T_2 / \Delta T_1)) \quad (12)$$

Utilizing this mean temperature, the following expression was derived:

$$Q_{\text{steam},n} = U_v \cdot V \cdot \Delta T_{1m} \quad (13)$$

Equation 13 was substituted into equation 9 to give

$$\begin{aligned} m_G (\bar{C}_{pO} |_{T_{n-1}} \cdot T_{n-1} - \bar{C}_{pO} |_{T_n} \cdot T_n) + \\ m_{w,n-1} \bar{C}_{pw} |_{T_{n-1}} \cdot T_{n-1} - m_{w,n} \bar{C}_{pw} |_{T_n} \cdot T_n - \\ Q_{\text{Loss},n} = U_v \cdot V \Delta T_{1m} \end{aligned} \quad (14)$$

Replacing the terms on the left side of equation 14 with $Q_{\text{total},n}$ and rearranging the terms yielded

$$U_v = Q_{\text{total},n} / (V \cdot \Delta T_{1m}) \quad (15)$$

Equation 15 was used to obtain the volumetric heat transfer coefficients desired.

Further calculations required for data correlation are given in Chapter V.

CHAPTER V

RESULTS AND DISCUSSION

The volumetric heat transfer coefficients and associated independent variables obtained using unmodified oil on Pall rings are given in Table I (p. 33). Each of these data sets is the average of data collected on all three stages under a given set of operating conditions. The same is true of the data collected using unmodified oil on Intalox[®] saddles, which are shown in Table II (p. 36). Table III (p. 38) includes the same information for data collected using modified oil on the Intalox[®] packing. Figure 10 (p. 40) illustrates the highest heat transfer obtained for each oil condition and packing type.

The largest volumetric heat transfer coefficients, using unmodified oil, were obtained with high oil and water flow rates and Intalox[®] saddles. Under these constraints, the highest volumetric heat transfer coefficients were obtained at low absolute pressure and high stagewise temperature differential.

The contribution of oil and water flow to the volumetric heat transfer was accounted for by the following terms:

TABLE I DATA: UNMODIFIED OIL-PALL RINGS

SATURATED STEAM 373 K: 2.22 K STAGewise ΔT

RUN NO.	U_v $J/m^3 \cdot s \cdot K$	OIL FLOW $kg/hr \cdot m^2$	WATER FLOW $kg/hr \cdot m^2$	STEAM FLOW $kg/hr \cdot m^2$	STEAM DENSITY kg/m^3	$\frac{L}{V}$	A_p
001PR	259 527	21 105	86.94	58.52	0.643 9	362.13	125
002PR	293 451	21 105	1 893.86	66.89	0.643 9	343.83	125
003PR	307 852	21 105	2 602.95	69.68	0.643 9	340.24	125
004PR	319 998	21 105	3 312.11	72.47	0.643 9	336.92	125
005PR	362 231	29 551	121.25	81.73	0.643 9	363.05	125
006PR	408 097	29 551	2 635.40	91.94	0.643 9	350.08	125
007PR	426 410	29 551	3 502.90	97.25	0.643 9	340.91	125
008PR	445 226	29 551	4 634.53	100.31	0.643 9	340.79	125
009PR	461 377	37 997	153.88	103.57	0.643 9	368.36	125
010PR	524 712	37 997	3 379.48	118.42	0.643 9	349.40	125
011PR	545 545	37 997	4 567.47	121.98	0.643 9	349.77	125
012PR	568 758	37 997	5 956.70	127.71	0.643 9	344.17	125

SATURATED STEAM 373 K: 1.67 K STAGewise ΔT

013PR	247 623	21 105	66.31	43.68	0.631 1	484.69	125
014PR	349 563	29 551	93.53	62.26	0.631 1	476.14	125
015PR	425 776	37 997	4 598.80	75.25	0.631 1	453.82	125
016PR	547 086	37 997	5 913.14	97.10	0.631 1	452.22	125

TABLE I--Continued

SATURATED STEAM 388 K: 1.11 K STAGewise ΔT

017PR	303 176	21 105	3 261.03	38.62	1.015 6	630.92	125
018PR	341 963	29 551	64.20	43.27	1.015 6	684.43	125
019PR	417 231	29 551	4 566.85	54.79	1.015 6	622.70	125
020PR	447 834	37 997	85.51	57.58	1.015 6	661.38	125
021PR	593 733	37 997	5 869.64	68.73	1.015 6	638.25	125

SATURATED STEAM 355 K: 3.33 K STAGewise ΔT

022PR	392 634	29 551	169.16	114.23	0.355 6	260.18	125
023PR	488 371	29 551	4 696.12	142.10	0.355 6	241.01	125
024PR	503 368	37 997	216.83	146.29	0.355 6	261.22	125
025PR	625 409	37 997	6 036.19	181.57	0.355 6	242.51	125

SATURATED STEAM 355 K: 2.22 K STAGewise ΔT

026PR	389 468	29 551	112.30	75.25	0.339 6	394.20	125
027PR	486 397	29 551	4 626.33	94.31	0.339 6	362.39	125
028PR	501 300	37 997	289.98	97.10	0.339 6	394.30	125
029PR	623 566	37 997	5 947.63	120.89	0.339 6	363.51	125

SATURATED STEAM 339 K: 3.33 K STAGewise ΔT

030PR	419 219	29 551	157.17	105.70	0.179 4	281.06	125
031PR	526 543	29 551	4 683.44	132.83	0.179 4	257.73	125
032PR	538 093	37 997	202.10	135.62	0.179 4	281.66	125
033PR	675 542	37 997	6 020.59	170.26	0.179 4	258.53	125

TABLE I--Continued

SATURATED STEAM 339 K: 4.44 K STAGewise ΔT

034PR	431 291	29 551	165.37	142.58	0.200 2	208.42	125
035PR	540 757	29 551	4 751.06	178.62	0.200 2	192.04	125
036PR	553 761	37 997	272.08	182.99	0.200 2	209.13	125
037PR	694 674	37 997	6 108.78	229.44	0.200 2	192.23	125

SATURATED STEAM 339 K: 5.55 K STAGewise ΔT

038PR	430 080	29 551	249.71	168.58	0.208 2	176.77	125
039PR	539 342	29 551	4 798.29	211.29	0.208 2	162.57	125
040PR	552 885	37 997	321.17	216.87	0.208 2	176.69	125

TABLE II DATA: UNMODIFIED OIL-INTALOX [®] SADDLES

SATURATED STEAM 339 K: 3.89 K STAGewise ΔT

RUN NO.	U_v J/m ³ .s.K	OIL FLOW kg/hr.m ²	WATER FLOW kg/hr.m ²	STEAM FLOW kg/hr.m ²	STEAM DENSITY kg/m ³	$\frac{L}{V}$	A_p
001IS	442 077	29 551	183.21	123.51	0.193 8	240.72	150
002IS	554 990	29 551	4 715.82	155.11	0.193 8	220.92	150
003IS	567 211	37 997	235.16	158.36	0.193 8	241.43	150
004IS	711 086	37 997	6 062.04	198.31	0.193 8	222.17	150

SATURATED STEAM 339 K: 5.55 K STAGewise ΔT

005IS	456 627	29 551	249.27	168.55	0.200 2	176.80	150
006IS	572 670	29 551	4 798.79	211.29	0.200 2	162.57	150
007IS	584 965	37 997	318.56	215.45	0.200 2	177.84	150

SATURATED STEAM 355 K: 3.33 K STAGewise ΔT

008IS	420 896	29 551	170.78	114.71	0.355 6	259.10	150
009IS	524 252	29 551	4 698.48	143.13	0.355 6	239.29	150
010IS	540 292	37 997	219.25	147.23	0.355 6	259.57	150
011IS	671 778	37 997	6 058.38	183.09	0.355 6	240.74	150

SATURATED STEAM 355 K: 2.39 K STAGewise ΔT

012IS	401 279	29 551	120.75	80.83	0.341 2	367.09	150
013IS	501 077	29 551	4 636.40	101.26	0.341 2	337.62	150
014IS	515 850	37 997	155.12	104.05	0.341 2	366.67	150
015IS	640 816	37 997	5 959.81	129.12	0.341 2	340.43	150

TABLE II--Continued

SATURATED STEAM 373 K: 2.22 K STAGewise ΔT

016IS	387 679	29 551	142.38	95.67	0.650 3	310.37	150
017IS	474 548	29 551	4 659.20	116.57	0.650 3	293.47	150
018IS	495 991	37 997	182.09	122.15	0.650 3	312.56	150
019IS	610 040	37 997	5 990.64	150.02	0.650 3	293.21	150

SATURATED STEAM 373 K: 1.67 K STAGewise ΔT

020IS	359 921	29 551	99.62	66.89	0.635 9	443.27	150
021IS	444 164	29 551	4 609.18	82.68	0.635 9	413.16	150
022IS	460 557	37 997	127.84	85.47	0.635 9	446.06	150
023IS	566 429	37 997	5 924.64	105.00	0.635 9	418.30	150

SATURATED STEAM 388 K: 1.11 K STAGewise ΔT

024IS	354 388	29 551	70.23	46.34	1.015 6	639.22	150
025IS	429 287	29 551	4 570.21	54.66	1.015 6	624.24	150
026IS	450 200	37 997	86.94	58.52	1.015 6	650.79	150
027IS	546 122	37 997	5 873.43	70.62	1.015 6	621.22	150

TABLE III DATA: AVERAGE MODIFIED OIL-INTALOX[®] SADDLES

SATURATED STEAM 373 K: 1.67 K STAGewise ΔT

RUN NO.	U_v J/m ³ .s.K	OIL FLOW kg/hr.m ²	WATER FLOW kg/hr.m ²	STEAM FLOW kg/hr.m ²	STEAM DENSITY kg/m ³	$\frac{L}{V}$	A_p
001MIS	427 714	29 551	91.16	61.31	0.643 3	483.48	150
002MIS	402 303	29 551	1 376.99	64.10	0.643 3	482.50	150

SATURATED STEAM 373 K: 1.88 K STAGewise ΔT

003MIS	409 867	29 551	2 678.33	77.10	0.643 3	434.94	150
004MIS	437 122	29 551	3 968.38	82.68	0.643 3	405.41	150

SATURATED STEAM 373 K: 1.67 K STAGewise ΔT

005MIS	546 030	37 997	116.55	78.05	0.643 3	488.32	150
006MIS	515 328	37 997	1 405.19	82.68	0.643 3	476.56	150

SATURATED STEAM 373 K: 1.88 K STAGewise ΔT

007MIS	512 981	37 997	2 708.00	96.62	0.643 3	421.29	150
008MIS	540 217	37 997	3 998.00	102.20	0.643 3	410.91	150
009MIS	564 566	37 997	5 285.70	106.84	0.643 3	405.12	150

TABLE III--Continued

SATURATED STEAM 339 K: 5.55 K STAGEWISE ΔT

010MIS	539 398	29 551	262.22	177.37	0.244 8	168.08	150
011MIS	692 147	37 997	336.64	227.98	0.244 8	168.14	150

SATURATED STEAM 339 K: 5.27 K STAGEWISE ΔT

012MIS	523 600	29 551	1 548.00	180.24	0.244 8	172.54	150
013MIS	663 470	37 997	1 619.42	226.25	0.244 8	173.41	150
014MIS	527 810	29 551	2 857.32	198.73	0.244 8	163.08	150
015MIS	656 819	37 997	2 926.40	245.64	0.244 8	166.60	150
016MIS	561 753	29 551	4 158.18	211.74	0.244 8	159.20	150
017MIS	692 457	37 997	4 292.27	259.11	0.244 8	163.21	150

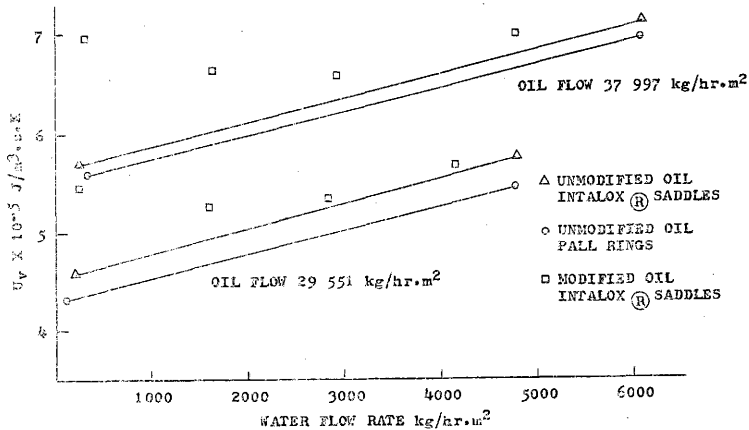


FIGURE 10. BEST VOLUMETRIC HEAT TRANSFER COEFFICIENTS

$$U_{v_f} = -0.0001003314 \cdot m_o \cdot m_w + 19.6923 \cdot m_w + 12.5026 \cdot m_o - 28549.73 \quad (16)$$

For the unmodified oil runs, term 16, liquid to vapor ratio, steam density, packing factor, and volumetric heat transfer coefficient were all correlated by use of a least squares linear regression program on an IBM 360/65 computer. The correlation obtained is given below:

$$U_v = C_{fit} \left(\frac{A_p^{0.2053} \cdot U_{v_f}^{0.9652}}{\rho_s^{0.0925} \cdot (L/V)^{0.0836}} \right) \quad (17)$$

The water to oil ratio was originally included as one of the independent variables. However, a very low power on the ratio and poor statistics were considered sufficient justification for omitting water to oil ratio in the final correlation. Equation 17 has an average error of 1.8 per cent with a maximum error of 7.9 per cent. The statistics on this curve-fit are given in Appendix III. It is notable that the correlation is dimensional and does not allow for variations in viscosity and surface phenomenon. Viscosity and surface phenomenon, in the unmodified oil runs, were not varied

sufficiently to be included in the curve-fit. Also, provision was not made to measure either of these parameters. Had these two factors been included, the data correlation might have been characterized in terms of a more common dimensionless group.

The stability of the physical character of the oil was in some question throughout the data collection. This was due to rust in the system and dissolved paint and light and heavy hydrocarbon impurities in the oil.

The power on U_{vf} in equation 17 in the data correlation supports the strong dependence of volumetric heat transfer on liquid flow suggested by Cheng (1963) and Sideman (1966).

The power on the packing factor, A_p , in equation 17, indicates that choosing a packing with a larger packing factor would improve heat transfer.

Steam flow was strongly dependent upon stagewise temperature differential, as was suggested by the work of Lackey (1961). Steam flow, alone, was not considered to be an important factor in the correlation since all of the previous work has indicated that volumetric heat transfer is independent of steam flow. However, steam flow is included in the liquid to vapor ratio. Liquid to vapor ratio and steam density were not strong contributors to the variation in heat transfer. Omission

of either of these in the data correlation procedure, however, produced poorer statistics and larger percentage error terms.

The number of data points collected using oil modified by 0.5 per cent octadecanol and 1.0 per cent octadecanol was not sufficient to make any curve-fit meaningful. The general trend associated with the addition of the "sensitizers" was to greatly improve, by as much as twenty per cent, heat transfer at low water to oil ratios. As the water to oil ratio was increased, the effect of the "sensitizer" became less pronounced. Another observation is that the variation of percentage of "sensitizer" used seemed to have no effect on the volumetric heat transfer coefficient. This may have been attributable to the fact that the oil was already partially "sensitized" by the presence of foreign materials. Visual inspection of the oil and pendant drop interfacial tension measurements tend to bear this out. This would also lead to the conclusion that truly pure oil would give poorer heat transfer than that attained using unmodified oil in this system. The gross effect of the addition of "sensitizers" to the oil was to increase heat transfer at the low water rate end of the overall system, such that the total

heat transfer on a full scale condenser would be greater than that obtained with unmodified oil.

The results of this work indicate that, while the values of U_v obtained are not as high as would have been desired, they are better than those obtained in the closest comparable work (Cheng 1963). The conclusions of Maa and Hickman (1972), with regard to the effect of "sensitizers", have been supported but require much further investigation.

The volumetric heat transfer coefficients obtained in this work tend to make direct-contact oil-steam condensation a possible contender, against other more conventional processes, on an industrial scale.

NOMENCLATURE

A_p	Packing factor	dimensionless
C	Constant	
\bar{C}_p	Mean heat capacity	Cal/gm.deg C
d	Particle diameter	
G_a	Galileo number	dimensionless
H	Enthalpy per unit mass	J/kg
HTU	Height of a transfer unit	ft
L/V	Liquid to vapor ratio	dimensionless
m	Mass flow rate	kg/hr
$Pr.$	Prandtl number	dimensionless
Q	Enthalpy	J
$Re.$	Reynolds number	dimensionless
S	Specific gravity	dimensionless
T	Temperature	
U_v	Volumetric heat transfer coefficient	$J/m^3.s.K$
V	Volume of packed section	m^3

GREEK SYMBOLS

\propto	Indicates proportionality
Δ	Specifies a change in the variable it precedes
λ	Latent heat J/kg
ρ	Density kg/m^3

NOMENCLATURE--Continued

SUBSCRIPTS

f	Flow
F	Fahrenheit
K	Kelvin
L	Liquid phase
lm	Log mean
n	Stage designation
o	Oil
®	Registered
s	Steam
w	Water
1	Initial condition
2	Final condition

SUPERSCRIPTS

d	Arbitrary exponent
f	Arbitrary exponent
h	Arbitrary exponent

BIBLIOGRAPHY

- Cheng, C. J., M.S. Thesis, University of California, Berkeley (1963).
- Drummond, J., Texas A&M University, private communication.
- Harriott, P., and H. Wiegandt, AIChE J., 10, 755 (1964).
- Kreith, F., Principles of Heat Transfer, 2nd Ed., pp. 490-492, International Textbook Company, Scranton, Pennsylvania (1969).
- Lackey, D. L., M.S. Thesis, University of California, Berkeley (1961).
- Maa, J. R., and K. Hickman, "Direct Condensation of Steam on a Modified Oil Coolant," Desalination, 10, 95 (1972).
- Perry, H., Chemical Engineers' Handbook, 4th Ed., p. 3-218, McGraw-Hill, New York (1971).
- Sideman, S., "Direct Contact Heat Transfer Between Immiscible Liquids," Advances in Chemical Engineering, 6, 207, Academic Press, New York (1956).
- Wilke, C. R., C. J. Cheng, V. L. Ledesma, and J. W. Porter, Chem. Eng. Progr., 59, 69 (1964).

APPENDIX I
CONVERSION FACTORS

The conversion factors in Table IV, below, are provided to make it easier to examine the results of this investigation in terms of units more commonly used in the field of Chemical Engineering.

TABLE IV
CONVERSION FACTORS

TO CONVERT TO	FROM	MULTIPLY BY
BTU/ft ³ .hr.deg F	J/m ³ .s.K	5.357 839 E-02
lbm/hr.ft ²	kg/hr.m ²	1.620 641 E-01
lbm/ft ³	kg/m ³	6.242 797 E-01
lbm/in ²	N/m ²	1.450 380 E-04
gal(US)	m ³	2.641 776 E+02
ft	m	3.280 833 E+00
deg F	K	T _F =1.8T _K -459.67

APPENDIX II

DATA USED IN HEAT LOSS CALCULATIONS

Prior to the data runs in which oil was used, it was necessary to determine the heat losses from the experimental system. These heat loss terms were then used in the calculation of the volumetric heat transfer coefficients.

The heat losses from the system were obtained in the following manner:

1. Steam was circulated counter-current with water throughout the condensation stages until the temperature of the water in the separation system reached 400 K.
2. Steam flow was then stopped and the condensation stages were allowed to completely flood with water.
3. The water circulation rate was then reduced to 45.35 kg/hr.
4. Since the desired temperature operation profiles were known, constant monitoring of the inlet temperature to each stage was maintained. When the inlet temperatures reached values that were expected during operation of the system with

oil, both the inlet and outlet temperatures for the stage in question were recorded.

5. This procedure was repeated three times in its entirety. The resulting temperatures were averaged and then used to determine the heat loss for each stage.
6. The stagewise heat losses were then curve-fit, encoded in Fortran IV, and used in the calculation of volumetric heat transfer coefficients.

Table V (p. 51) shows the average temperature differentials obtained by the above procedure.

TABLE V
TEMPERATURE DIFFERENTIALS USED IN
HEAT LOSS CALCULATIONS

STAGE 1		STAGE 2		STAGE 3	
TEMPERATURE (K)		TEMPERATURE (K)		TEMPERATURE (K)	
IN	OUT	IN	OUT	IN	OUT
388.50	387.78	389.61	388.77	390.72	389.84
373.00	372.33	375.22	374.50	376.88	376.10
352.50	352.25	356.28	355.67	359.06	358.39
338.60	338.10	343.04	342.49	347.48	346.87

APPENDIX III

DATA CORRELATION AND STATISTICS

Table VI (p. 53) contains the volumetric heat transfer coefficients obtained from enthalpy balance calculations on the experimental system, the volumetric heat transfer coefficients predicted by the correlation in equation 17, the difference between these two values, and the percentage error terms. Table VII (p. 55) shows the pertinent statistics associated with the data correlation.

TABLE VI
EXPERIMENTAL AND CORRELATED
VOLUMETRIC HEAT TRANSFER COEFFICIENTS

RUN NO.	EXP. U_v $J/m^3 \cdot s \cdot K$	FITTED U_v $J/m^3 \cdot s \cdot K$	DIFF. $J/m^3 \cdot s \cdot K$	PER CENT DIFF.
001PR	259 527	257 786	- 1 741	-0.671
002PR	293 451	292 334	- 1 117	-0.381
003PR	307 852	305 683	- 2 169	-0.704
004PR	319 998	319 017	- 981	-0.307
005PR	362 231	368 400	6 169	1.703
006PR	408 097	413 169	5 072	1.243
007PR	426 410	430 874	4 464	1.047
008PR	445 226	448 764	3 538	0.795
009PR	461 377	477 200	15 823	3.429
010PR	524 717	531 991	7 274	1.386
011PR	545 545	552 927	7 382	1.353
012PR	568 757	574 673	5 916	1.040
013PR	247 623	251 676	4 053	1.637
014PR	349 563	360 345	10 782	3.084
015PR	425 776	438 352	12 585	2.956
016PR	547 986	562 058	14 072	2.568
017PR	303 176	279 177	- 23 999	-7.916
018PR	341 963	334 067	- 7 896	-2.309
019PR	417 281	408 032	- 9 249	-2.217
020PR	447 834	434 642	- 13 192	-2.946
021PR	539 733	521 946	- 17 787	-3.296
020PR	392 634	401 102	8 468	2.157
023PR	428 371	489 201	830	0.170
024PR	503 368	519 958	16 590	3.296
025PR	625 409	626 565	1 156	0.185
026PR	389 486	388 030	- 1 456	-0.374
027PR	486 397	473 549	- 12 848	-2.641
028PR	501 300	505 775	4 475	0.893
029PR	623 566	606 747	- 16 819	-2.697
030PR	419 219	424 331	5 112	1.219
031PR	526 543	518 008	- 8 535	-1.621
032PR	538 093	550 192	12 099	2.248
033PR	675 542	663 668	- 11 874	-1.758

TABLE VI--Continued

034PR	431 291	430 835	-	456	-0.106
035PR	540 757	526 903	- 13	854	-2.526
036PR	553 761	559 682	-	5 921	1.069
037PR	694 674	675 106	- 19	568	-2.817
038PR	430 080	436 947	-	6 867	1.597
039PR	539 341	533 319	- 6	022	-1.117
040PR	552 885	566 509	- 13	624	2.464
001IS	442 077	443 617	-	1 540	0.348
002IS	554 990	541 522	- 13	468	-2.427
003IS	567 211	575 101	-	7 890	1.391
004IS	711 086	693 595	- 17	491	-2.460
005IS	456 627	453 595	-	3 032	-0.664
006IS	572 670	553 666	- 19	004	-3.318
007IS	584 965	587 742	-	2 777	0.475
008IS	420 896	416 574	- 4	322	-1.027
009IS	524 252	508 205	- 16	047	-3.061
010IS	540 292	540 119	-	173	-0.032
011IS	671 778	651 264	- 20	514	-3.054
012IS	401 279	405 220	-	3 941	0.982
013IS	501 077	494 594	- 6	573	-1.312
014IS	515 850	525 609	-	9 759	1.892
015IS	640 816	633 337	- 7	479	-1.167
016IS	387 679	387 525	-	154	-0.040
017IS	474 548	471 752	-	2 796	-0.589
018IS	495 991	502 267	-	6 276	1.265
019IS	610 040	604 656	- 5	384	-0.883
020IS	359 921	376 173	-	16 252	4.516
021IS	444 164	458 523	-	14 359	3.233
022IS	460 557	487 667	-	27 110	5.886
023IS	566 429	587 082	-	20 653	3.646
024IS	354 388	348 837	-	5 501	-1.552
025IS	429 297	424 280	-	5 017	-1.169
026IS	450 200	451 851	-	1 651	0.367
027IS	546 123	543 133	-	2 990	-0.548

TABLE VII
CURVE-FIT STATISTICS

t-TEST

MODEL: $\ln U_V = B_1 \ln A_p + B_2 \ln e_s + B_3 \ln L/V + B_4 \ln U_{V_f} + B_0$

HYPOTHESIS: Parameter has no significance in curve-fit.

DEGREES OF FREEDOM: 66 P: 0.1 $t_{P,66} = 1.645$

PARAMETER	COEFFICIENTS	t-VALUES
B_1	0.2053	6.199
B_2	-0.0925	-9.528
B_3	-0.0836	-5.618
B_4	0.9652	66.626

$$C_{fit} = e^{B_0}$$

$$B_0 = -0.02379$$

U_V is determined by evaluating the antilog of the MODEL. The independent variables were chosen by use of the orthogonal array method.

VITA

Christopher Alexander Wain was born on March 7, 1949 to 2nd/Lt. and Mrs. Weldon H. Wain in Lubbock, Texas. His elementary and secondary education was spread over fourteen different school districts in California and Texas. He graduated from Clear Creek High School in League City, Texas in 1967.

Christopher graduated from Texas A&M University in 1971 with a Bachelor of Science Degree in Chemical Engineering. His professional experience includes two years as an engineering design technician at the Texas A&M University Cyclotron Institute.

At the present time, Christopher is serving in the United States Air Force, and any future correspondence may be addressed in care of 1117 Scenic Loop, Marshall, Texas 75670.

The typist for this thesis was the author's wife, Janet Wain.

