**ChE laboratory**

# TWO EXPERIMENTS FOR ESTIMATING FREE CONVECTION AND RADIATION HEAT TRANSFER COEFFICIENTS

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THIS ARTICLE DESCRIBES two simple undergraduate heat transfer experiments which, when properly understood, may reinforce a student's understanding of free convection and radiation.

The purposes of the experiments are:

- **To demonstrate how the combined individual coefficients for free convection and radiation may be extracted from an experimentally determined overall heat transfer coefficient by arranging experimental conditions in such a way that the major resistances to heat transfer are the two of interest and all other resistances are so small as to be negligible.**
- **To illustrate a technique for reducing the radiation heat transfer coefficient to such a small value that a close estimation of the free convection coefficient is possible.**

In designing experiments for our undergraduates, we have attempted to keep them as simple as possible, and we wish to have the principle or principles they demonstrate to be so obvious that they are difficult to overlook. Also, we wish to have the results agree fairly closely with those values reported in the standard texts or with those values which would be calculated from standard correlations. Finally, we believe that students should either do or see done experiments which demonstrate the major phenomena we cover in our lecture courses. Although both free convection and radiation are covered in most lecture courses on heat transfer, few simple laboratory experiments seem to be available. We hope that the two experiments about to be described will be of some interest and use to instructors and students concerned with this subject.

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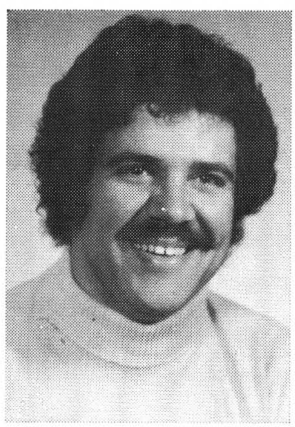
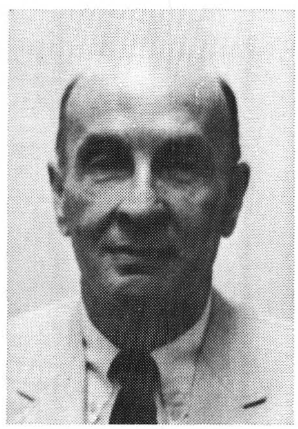
## APPARATUS

THE FUNCTION OF the apparatus is to provide data from which one may calculate an overall heat transfer coefficient. It consists, in its basic form, of a vertical glass tube exposed to the air, into the bottom of which is passed a stream of saturated vapor. As the vapor rises up through the tube and passes out the top of it, a portion





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**James O. Maloney** is professor of Chemical Engineering at the University of Kansas. He received his B.S. Degree from the University of Illinois and his M.S. and Ph.D. Degrees from the Pennsylvania State University. He has been employed by the Dupont Company and has held Fulbright posts at Naples, Italy; Cairo, Egypt; and Athens, Greece. (R) of the vapor condenses on the walls and runs down to the bottom of the tube. This condensate is drawn off and its rate of production measured. Knowing this, the latent heat of condensation, the area and the temperature difference, one may calculate the heat loss from the tube and the overall heat transfer coefficient.

Thus:

$$(\dot{m})(h_v) = Q/\theta = U(\pi D_o L)(T_V - T_A)$$

where

$\dot{m}$ = lb. condensate per hour
$h_v$ = latent heat of condensation, Btu/lb.
$Q/\theta$ = heat loss from the tube, Btu/hr.
$D_o$ = outside diameter of the tube, ft.
$L$ = length of the tube on which condensation occurs, ft.
$T_V$ = temperature of the saturated vapor, °F.
$T_A$ = temperature of the air, °F.
$U$ = overall heat transfer coefficient, Btu/(hr. ft² °F)

When this overall coefficient is determined experimentally, it is found to range from 2.2 to 3.4.

A consideration of this heat transfer process shows that there are three resistances in series to heat transfer: that resulting from the condensing vapor, that from the glass tube, and that of the combined resistance of free convection and radiation from the outside surface of the tube. That is:

$$R_{total} = R_{(cond.\ vapor)} + R_{tube} + R_{(conv.\ +\ rad.)}$$

$$= \frac{1}{(hA)_{\substack{cond.\ vapor \\ inside\ surface}}} + \left(\frac{L}{kA}\right)_{tube} + \left[\frac{1}{(h_{conv.} + h_{rad.})A}\right]_{\substack{outside \\ surface}}$$

The resistance of the condensing vapor is in the range of 0.005 to 0.0005 reciprocal Btu/(hr. ft.² °F), i.e., coefficients of 200 to 2,000. The resistance of the glass tube is about 0.013 [2.5 mm. thick, borosilica glass and a k of 0.63 Btu/(hr. ft. °F)]. Free convection coefficients on the outside of the tube are in the range of 1 to 2 Btu/hr. ft.² °F), and the radiation coefficients are in the same range or less. One can thus draw the preliminary conclusion that the overall coefficient will be essentially composed of the combination of a convection and a radiation coefficient from the outside of the tube.

Glass has an emissivity of about 0.9, but if it is wrapped tightly with polished aluminum foil, the emissivity of the vertical tube would be reduced from about 0.9 to about 0.05. Thus by wrapping the tube with aluminum foil the radiation coefficient can be so markedly reduced that the overall coefficient essentially equals the free convection coefficient.

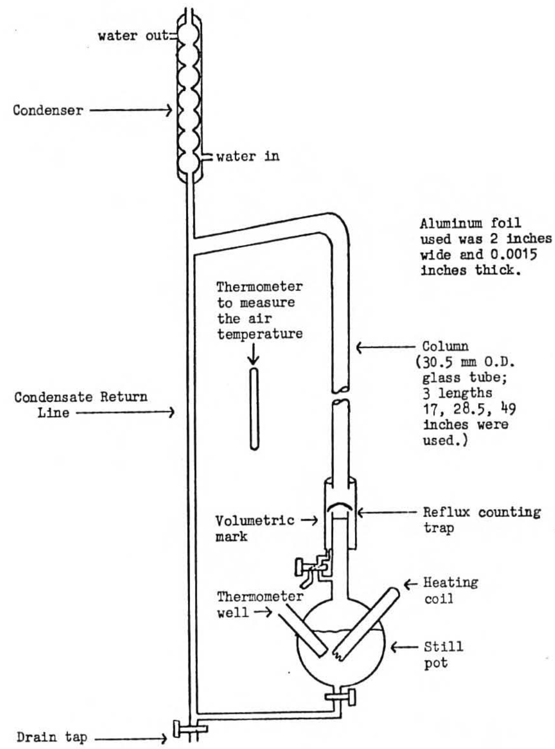


**FIGURE 1. Vertical Column for Heat Transfer Coefficient Measurements**

The apparatus actually used in the experiment is shown in schematic form in Figure 1. The only unusual feature was the reflux counting trap [1] which was a vacuum jacketed device that allowed measurement of a known volume of liquid condensate and separated the condensate stream from the rising vapor stream. Any device which will accomplish this purpose would be satisfactory. The excess vapor, which is necessary in order to be certain that the entire length of the column is at a constant temperature, does not need to be recycled provided the still pot contains enough liquid for several runs. The boil-up rate had no significant effect on the overall heat transfer coefficient as long as excess vapor left the top of the tube. The liquid in the still pot was vaporized by an immersion-type heater, but an exterior heater would, of course, work satisfactorily. A certain amount of care probably should be used to minimize the effects of any forced convection around the column. In our experiments, the air condition-

**TABLE 1. Effect of the Tube Height and Surface Condition on Overall Heat Transfer Coefficients**

| Tube Height in. | Column Surface Condition | Temperature, °F. n-Butanol | Temperature, °F. Air | Condensate Rate grams/hr. | $\dot{Q}$ Btu/hr. | U Btu/(hr.)(sq.ft)(°F.) |
|---|---|---|---|---|---|---|
| System: n-Butanol | | | | | | |
| 17 | Bare Glass | 242.6 | 82.4 | 383.9 | 215.3 | 3.02 |
| 28.5 | Bare Glass | 242.6 | 75.2 | 680.8 | 381.7 | 3.05 |
| 49 | Bare Glass | 242.6 | 82.4 | 1149.0 | 644.2 | 3.13 |
| 17 | Aluminum Wrapped | 241.7 | 76.1 | 227.4 | 127.5 | 1.73 |
| 28.5 | Aluminum Wrapped | 242.6 | 75.2 | 380.7 | 213.5 | 1.71 |
| 49 | Aluminum Wrapped | 242.6 | 82.4 | 624.0 | 349.9 | 1.70 |
| System: Water | | | | | | |
| 28.5 | Bare Glass | 211.1 | 73.4 | 129.0 | 276 | 2.69 |
| 28.5 | Aluminum Wrapped | 211.1 | 73.4 | 76.3 | 163 | 1.57 |

Example: First Run

$$\text{Area} = \pi D_0 L = (\pi)\left(\frac{30.5}{10} \times \frac{1}{2.54}\right)\left(\frac{17.0}{1.0}\right)\left(\frac{1}{144}\right) = 0.445 \text{ sq.ft.}$$

$$U = (215.3)/((242.6 - 82.4)(0.445))$$

$$= 3.02 \text{ Btu/(hr.)(sq.ft.)(°F.)}$$

ing vents were covered, and the windows were closed. In several runs, the entire assembly was surrounded with a vertical cardboard shield about two feet away from the column, but no significant difference in results was found. The columns we used were fitted with TS joints for ease of assembly.

## EXPERIMENTAL PROCEDURE

Two main types of runs were done: (1) those in which the glass tube was left bare, and (2) those in which the glass tube was covered with aluminum foil. During either type of run a reasonably pure liquid was placed in the still pot where it was boiled. Vapor ascended through the reflux trap into the tube. A certain amount condensed on the sides of the column, and the remaining vapor passed through the opening on the top of the column and off into the adjoining condenser. The condensate from it was then recycled via a condensate return and back into the still pot.

Whenever a measurement was desired, the three-way valve attached to the reflux trap was closed. The time required for the reflux trap to fill up to a marked level was recorded. Then the condensate was allowed to flow back into the still pot. This measuring procedure was repeated several times. After the last timing, the condensate in the reflux trap (up to the marked level) was collected and weighed.

During the run the temperature of the boiling liquid was measured by a thermometer immersed in a well in the still pot. In some later runs a thermometer was placed in the top of the tube to measure the vapor temperature there. The still pot temperature and this temperature were essentially identical. The ambient temperature was also measured.

A brief study was made of the effect of tube length on the overall heat transfer coefficients. Tubes of 17, 28.5 and 49 in. in length were used.

**TABLE 2. Effect of Temperature Difference or the Overall Heat Transfer Coefficient**

| Compound | Difference in Temperature Between Boiling Liquid and Air °F. | Condensate Rate grams/hr. | $\dot{Q}$ Btu/hr. | U Btu/(hr.)(sq.ft.)(°F.) |
|---|---|---|---|---|
| Acetone | 59.9 | 132.2 | 63.5 | 1.42 |
| Methanol | 66.6 | 71.1 | 80.2 | 1.61 |
| Benzene | 96.5 | 308.2 | 115.2 | 1.57 |
| Water | 137.7 | 73.4 | 163.0 | 1.57 |
| Toluene | 155.3 | 596.7 | 201.5 | 1.74 |
| n-Butanol | 161.8 | 390.8 | 219.4 | 1.82 |
| n-Octanol | 307.6 | 1549.7 | 599.6 | 2.61 |

Example: First Run

$$\text{Area} = \pi D L = (\pi)\left(\frac{30.5}{10.0} \times \frac{1}{2.54}\right)\left(\frac{28.5}{1}\right)\frac{1}{144}$$

$$= 0.747 \text{ sq. ft.}$$

$$U = \frac{(Q/\theta)}{(A)(\Delta T)} = \frac{63.5}{(0.747)(59.9)} = 1.42 \text{ Btu/(hr.)(sq.ft.)(°F.)}$$

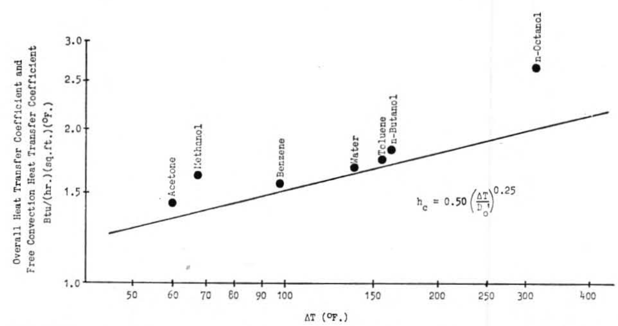


**FIGURE 2. Measured Overall Heat Transfer Coefficients Together with a Plot of the Perry Third Edition Formula for Free Convection Coefficients**

In order to have a variety of temperature driving forces, a number of liquids of various boiling points were used, including: acetone, benzene, toluene, water, methanol, n-butanol, n-amyl alcohol, and n-octanol (BP, range 56-195°C.).

## TYPICAL RESULTS

SOME TYPICAL RESULTS are shown in Tables 1 and 2. Table 1 shows two things. It shows that the overall coefficient is essentially independent of tube heights from 17 to 49 inches and that wrapping the tube tightly with aluminum foil results in a marked decrease in the overall coefficient from 3.1 to 1.7. Table 2 and Figure 2 show that the overall coefficient generally increases as the temperature difference between the boiling points of the compound and the air increases.

## DISCUSSION

IT IS CUSTOMARY to have students compare their results calculated from experimental data with values found in the literature. The procedure used here is to compare the experimental value with ones calculated using the methods found in the five consecutive editions of Perry's Chemical Engineers' Handbook. The particular comparison shown here is for the case of water condensing at 735 mm. Hg and 99.5°C. inside a 28.5 inch long vertical tube 1.2 inches in outside diameter. The tube was carefully wrapped with shiny aluminum foil. The air temperature was 23°C. The measured overall U was 1.57 Btu/(hr. ft.$^2$ °F.).

The principal reason for selecting the procedures found in Perry is that it is a reference more widely available to undergraduates than any particular text on heat transfer.

While all of the details of calculating the values by each method will not be gone through, the values used in the calculation will be shown.

First Edition [2]:

$$h_{\text{vert. cylinder}} = h_{\text{hor. cylinder}} \text{ x shape factor}$$

$$h_{\text{hor. cylinder}} = f\left(\text{film temp.}, \frac{P^2 \Delta T_m}{D}\right)$$

$$\text{film temp.} = \frac{(99.5 + 23)}{2} = 61.3°C$$

$$\frac{P^2 \Delta T_m}{D} = \frac{(0.97)^2 (99.5 - 23)}{1.2} = 60$$

$$h_{\text{hor. cylinder}} = 1.2 \text{ Btu/(hr. ft.}^2\text{ °F.)}$$

$$\text{shape factor} = 1.22$$

$$h_{\text{vert. cylinder}} = (1.2)(1.22)$$

$$= 1.46 \text{ Btu/(hr. ft.}^2\text{ °F.)}$$

Second Edition [3]:

For long vertical pipes:

$$h_c = 0.4 \left(\frac{\Delta t_s}{D_o'}\right)^{0.25}$$

$$\Delta t_s = 211.1 - 73.4 = 137.7$$

$$h_c = 0.4 \left(\frac{137.7}{1.2}\right)^{0.25} = 1.31 \text{ Btu/(hr. ft.}^2\text{ °F.)}$$

Third Edition [4]:

For long vertical pipes:

$$h_c = 0.5 \left(\frac{\Delta t_s}{D_o'}\right)^{0.25}$$

$$= 0.5 \left(\frac{137.7}{1.2}\right)^{0.25} = 1.64 \text{ Btu/(hr. ft.}^2\text{ °F.)}$$

Fourth Edition [5]:

For vertical surfaces:

$$\frac{h_c L}{k} = a(x)^m$$

for $x > 10^9$; $a = 0.13$, $m = 1/3$

$$x = \left[\frac{L^3 \rho^2 g \beta \Delta t}{\mu^2} \quad \frac{c\mu}{k}\right]_{\text{film temperature}}$$

$$x = \frac{(2.38)^3 (0.0636)^2 (4.18\text{x}10^8) \; (0.001673)(137.7)(0.70)}{(0.0465)^2}$$

$$= 1.7\text{x}10^9$$

$$h_c = (0.13)(1.7\text{x}10^9)^{1/3} \left(\frac{0.0166}{2.38}\right)$$

$$= 1.08 \text{ Btu/(hr. ft.}^2\text{ °F.)}$$

Fifth Edition [6]:

For vertical surfaces:

$$N_{N\mu} = a(N_{Gr} N_{Pr})^m$$

for $x > 10^9$; $a = 0.13$, $m = 1/3$

This is the same procedure as found in the fourth edition, so:

$$h_c = 1.08 \text{ Btu/(hr. ft.}^2\text{ °F.)}$$

The results of these calculations are summarized in Table 3. One sees considerable variation among the results. Two comments may be made about the five procedures. Nothing is stated about how well these calculated values agree with experimental values. Furthermore, as the calculational procedures change from edition to edition, no reasons are stated for making the changes. A student looking at the wide range of these values might be somewhat confused or not know exactly how to proceed; without access to the actual data upon which the methods in Perry are based, (s)he is probably at an impasse.

**TABLE 3**
**Natural Convection Coefficients for Air As Calculated from Procedures Found in Five Consecutive Editions of Perry's *Chemical Engineers' Handbook***

| Edition | Date | $h_c$, Btu/(hr. ft.$^2$ °F.) |
|---|---|---|
| First | 1934 | 1.46 |
| Second | 1941 | 1.31 |
| Third | 1950 | 1.67 |
| Fourth | 1963 | 1.08 |
| Fifth | 1973 | 1.08 |

Table 4 and Figure 2 have been prepared using experimental results and the correlation

**TABLE 4. Comparison of Overall Measured U with the Calculated Free Convection Coefficients for the Aluminum Wrapped Tube**

| Substance | U Measured Btu/(hr.)(sq.ft.)(°F.) | $h_c$ Calculated Btu/(hr.)(sq.ft.)(°F.) |
|---|---|---|
| Acetone | 1.42 | 1.34 |
| Methanol | 1.61 | 1.37 |
| Benzene | 1.57 | 1.51 |
| Water | 1.57 | 1.67 |
| Toluene | 1.74 | 1.70 |
| n-Butanol | 1.82 | 1.71 |
| n-Octanol | 2.61 | 2.01 |

equation from the third edition of Perry:

$$h_c = 0.5\left(\frac{\Delta t_s}{D_o'}\right)^{0.25}$$

$h_c$ = free convection coefficient, Btu/(hr. ft.$^2$ °F.)
$\Delta t_s$ = temperature difference between the hot surface and the ambient air, °F.
$D_o'$ = diameter of the tube, in.

One observes that the trend of the values is similar to that predicted by the equation and that $h_c$ is usually lower than the overall heat transfer coefficient, as would be expected.

Another calculation can be made to take the radiation heat loss into account. The radiation heat loss can be estimated from Stefan's Law:

$$\dot{Q}_r = \sigma \epsilon A (T_s^4 - T_a^4)$$

where

$\dot{Q}_r$ = heat transfer rate, Btu/hr.
$\sigma$ = $0.1717 \times 10^{-8}$
$\epsilon$ = tube surface emissivity
A = the area of heat transfer, sq. ft.
$T_s$ = surface temperature of the tube, °R.
$T_a$ = air temperature, °R.

Table 5 shows the results of two calculations in which the amount of heat transferred by radiation is estimated from the 28.5 in. long tube, using n-butanol as the boiling material. In the first case, the tube is bare and the glass emissivity is taken as 0.9. In the second case, the aluminum emissivity is taken as 0.05. The two emissivity values are taken from Kreith [7]. From the heat lost by radiation, the two radiation heat transfer coefficients were calculated. One sees from the calculation that an aluminum covering should markedly reduce the coefficient, which is exactly what happened experimentally.

**TABLE 5. Estimated Heat Loss by Radiation and the Radiation Heat Transfer Coefficient**

| Tube Condition | Emissivity, $\epsilon$ | $\dot{Q}_{total}$ Btu/hr. | $\dot{Q}_r$ Btu/hr. | $\dot{Q}_{free\ convection}$ Btu/hr. | $h_r$ |
|---|---|---|---|---|---|
| Bare Glass | 0.9 | 381.7 | 185.9 | 195.8 | 1.47 |
| Aluminum Covered | 0.05 | 213.5 | 10.3 | 205.2 | 0.08 |

The two values of $h_r$ calculated may be used together with the measured overall U for the bare tube to estimate the overall coefficient for the aluminum foil wrapped tube as follows:

| | |
|---|---|
| overall U with bare tube | 3.05 Btu/(hr. ft.$^2$ °F.) |
| reduction by radiation coefficient of bare tube | 1.47 Btu/(hr. ft.$^2$ °F.) |
| net $h_c$ | 1.58 Btu/(hr. ft.$^2$ °F.) |
| add back radiation coefficient of Al-foil surface | 0.08 Btu/(hr. ft.$^2$ °F.) |
| estimated overall U, Al-wrapped tube | 1.66 Btu/(hr. ft.$^2$ °F.) |
| measured overall U, Al-wrapped tube | 1.71 Btu/(hr. ft.$^2$ °F.) |

One sees that the estimated and measured overall U for this case are very close.

## CONCLUSIONS

- **These experiments provide a means for measuring overall heat transfer coefficients under conditions where free convection and radiation are controlling the transfer.**
- **The experiments show how the radiation coefficient may be reduced so much that free convection is controlling the heat transfer.**
- **Application of procedures found in successive editions of Perry's *Chemical Engineers' Handbook* to estimate free convection coefficients give significantly different answers.**
- **A closer inspection of the data on which the correlating procedures found in Perry are based seems warranted.** □

## REFERENCES


1. Scientific Glass and Instruments, Inc., Catalog, Houston, Texas, Item 9190.
2. Perry, J. H., *Chemical Engineers' Handbook*, McGraw-Hill, 1st Ed., 1934, pp. 861-864.
3. *Ibid*, 2nd Ed., 1941, pp. 985-987.
4. *Ibid*, 3rd Ed., 1950, pp. 474-476.
5. *Ibid.*, 4th Ed., 1965, pp. 10-10 — 10-13.
6. *Ibid*, 5th Ed., 1973, pp. 10-10 — 10-12.
7. Kreith, Frank, *Principles of Heat Transfer*, 3rd Edition, Intext Press, Inc., 1973, pp. 236-237.