Chapter V

DISCUSSIONS

5.1 Some Considerations on the Design of Experimental Apparatus

The main interest in studying a vertical film evaporator has been the determination of film heat transfer coefficient. Essential parts of the experimental apparatus were carefully chosen according to their functions. The preheater was required because, in a film evaporator, the feed had to enter the evaporator at a temperature very close to its boiling point. Falling film evaporators are usually designed with built-in preheaters.

Steam entrainment seperator was included to improve steam quality.

A distributor was designed to be mounted on the preheater cover as shown in Fig 3.4. Accurate adjustment of its position was required to ensure uniform film of desired thickness on the evaporating surface.

To measure vapor temperature and wall temperature, thermocouples were used. For vapor phase temperature, a thermocouple was inserted in a small glass tube, and the tube was then centrally inserted from the top to the mid-section of the heated tube. By so doing, the end of the

thermocouple was not in contact with the liquid film and the recorded temperature was a true vapor phase temperature. For the measurement of wall temperature, a hole was drilled through the middle of the outer tube wall and a small stainless steel tube was pushed through the hole against the inner tube wall. A thermocouple was inserted and had its junction touched the wall of the inner tube. The wall of this tube was very thin that the thermocouple was not welded on to it.

5.2 Variations of h_m with Operating Parameters

5.2.1 Effect of Feed Rate on h_m

The effect of feed rate on h_m was illustrated in Fig 4.5. Both feed rate and h_m were expressed in terms of Re and $h_m \left(\dfrac{\mu^2}{\rho^2 gk^3} \right)^{1/3}$. The calculated values were shown in Table D-3. It was observed that h_m decreases up to Re of 1.2×10^3 and then continue to increase with. Re. It could be explained by the mechanism of transfer in laminar and turbulent flow. For Re below 1.2×10^3, fluid flows in the laminar regime, heat transfers by molecular transport from the heated surface to liquid-vapor interface. Increasing Re would increase film thickness, resulting in the increase of the resistance to heat transfer; thus the heat transfer coefficient was lower. For turbulent regime, Re beyond

1.2×10^3, eddy motion decreased laminar sublayer thickness, heat could transferr at a higher rate than in laminar case. Increasing Re resulted in the increasing of turbulence, heat transfer coefficient was therefore higher. From Re of 2.0×10^3, h_m sharply drops again. This was dued to over-shooting of the feed rate at the distributor, sothat not the whole of the liquid attached itself to the entire tube sur-face as a falling film. Part of the liquid dropped down through the cavity of the tube, thus the heat transferr was lower.

The experimental h_m were plotted as points in Fig 5.1 to compare with predicted values of heat transfer coefficient for the Nusselt type and for turbulence given by Dukler, Chun and Seban. It was observed that for laminar regime, the experimental results agreed well with Nusselt's equation. For turbulent regime with Pr = 1.77 the h_m value varied with Re as it was shown in a form of Chun and Seban expression. Since the maximum feed rate in the experiments of Dukler, Chun and Seban were not beyond the overshooting limit of their apparatus, the sharply dropped line due to above reason were not appeared.

5.2.2 Effect of the Wall Temperature on h_m

The effect of the wall temperature on h_m is illus-trated in Fig 4.7. Steam inlet temperature was used in
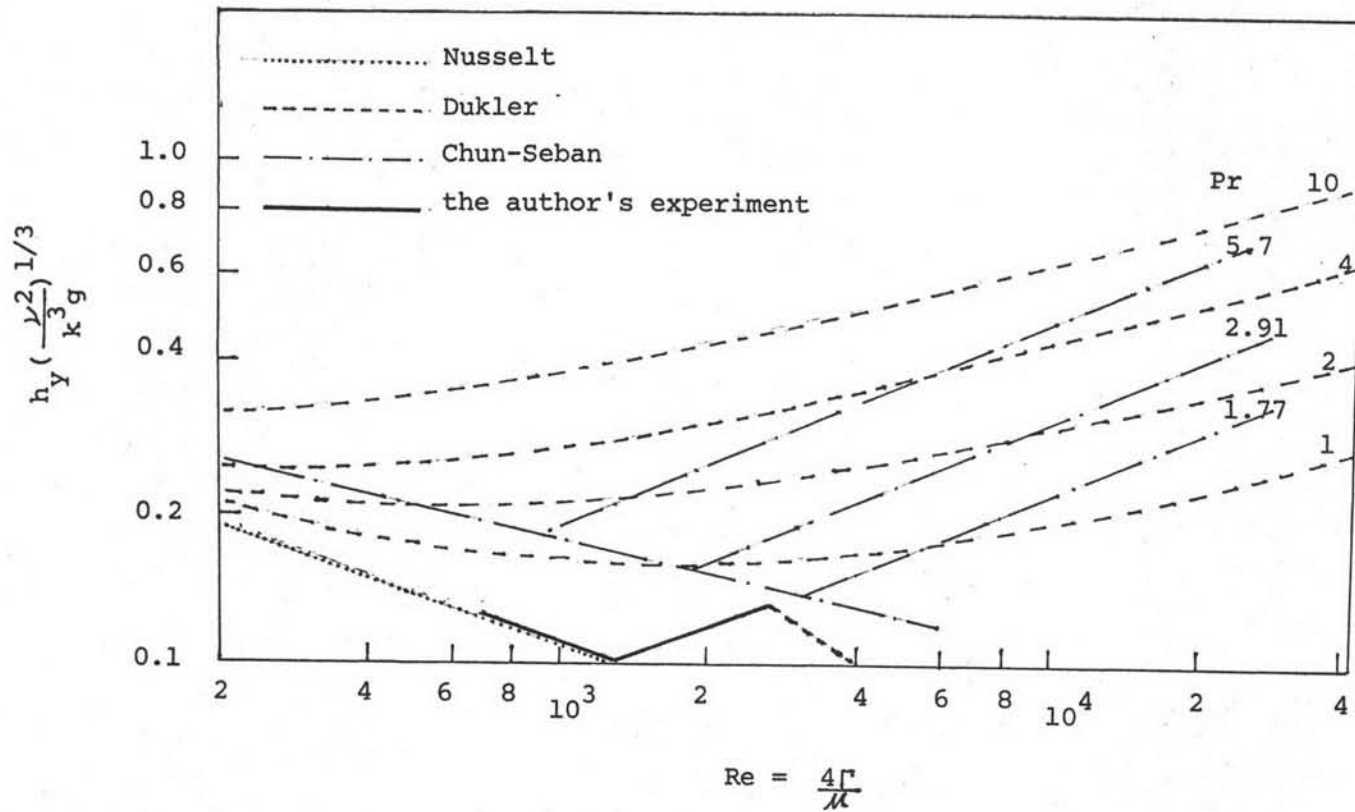
Fig.5.1 Local heat transfer coefficient as a function of Reynolds
number, comparison of theory with the author's experiment.

place of the wall temperature. In the evaporating tube, liquid was heated until its temperature reached 212°F and then it evaporated. Consequently, h_m was the combined heat transfer coefficient of preheating zone and evaporating zone. At the steam temperature of 235°F., only small amount of water evaporated, heat transfer coefficient was calculated from the total heat transferred which was mainly the sensible heat of the liquid. Because the resistance heat transfer, in case of preheating was the wall resistance only, heat transfer coefficient was rather high. By increasing steam temperature larger quantity of water was evaporated, heat transfer coefficient was calculated from the total heat transferred which was the latent heat of evaporation. The heat transfer coefficient, h_m was lower than when the liquid was preheated.

As steam temperature increased above 259°F, formation of bubble nuclei increased, and the increase in bubble activity produced turbulence near the surface of the film. Consequently, the heat transfer coefficient was increased. However, the heat transfer coefficient appeared to be a constant above the steam temperature 280°F.

To illustrate the effect of temperature difference $T_s - T_v$, clearly, the evaporated fraction and the total heat transfer calculated from experimental results of a 2-ft tube and feed temperature of 198°F were plotted

against the temperature differences as shown in Fig 5.2. By increasing the temperature difference, heat transfer was increased. For a Re of 2.0×10^3 when fluid flowed in turbulent regine, heat could be transferred at a higher rate than the case of laminar flow, i.e. Re of 1.3×10^3.

As a result of the increase in heat transfer rate, the water evaporated at Re of 2.0×10^3 is greater than that at Re of 1.3×10^3. Nevertheless, the ratio of the water evaporated to the original feed, i.e. evaporated fraction, for Re of 2.0×10^3 is less than that for Re of 1.3×10^3. The difference is shown in Fig 5.2.

5.2.3 Effect of Viscosity on h_m

Effect of viscosity on h_m is shown in Fig 4.9 by plotting h_m with feed rate at different sucrose concentration. To illustrate the effect clearly, h_m were read from Fig 4.9 at feed rate of 250, 350 and 450 lb/hr ft and plottedagainst concentration of sucrose solutions as shown in Fig 5.3. h_m decreased when viscosity (concentration) increased. This is due to the increasing of boundary layer with viscosity.

When h_m is replotted in term of $h_m \left(\dfrac{\mu^2}{\rho^2 g k^3} \right)^{1/3}$ against Reynolds number calculated with physical properties at $212^{\circ}F$, the variation of h_m with feed rate at various
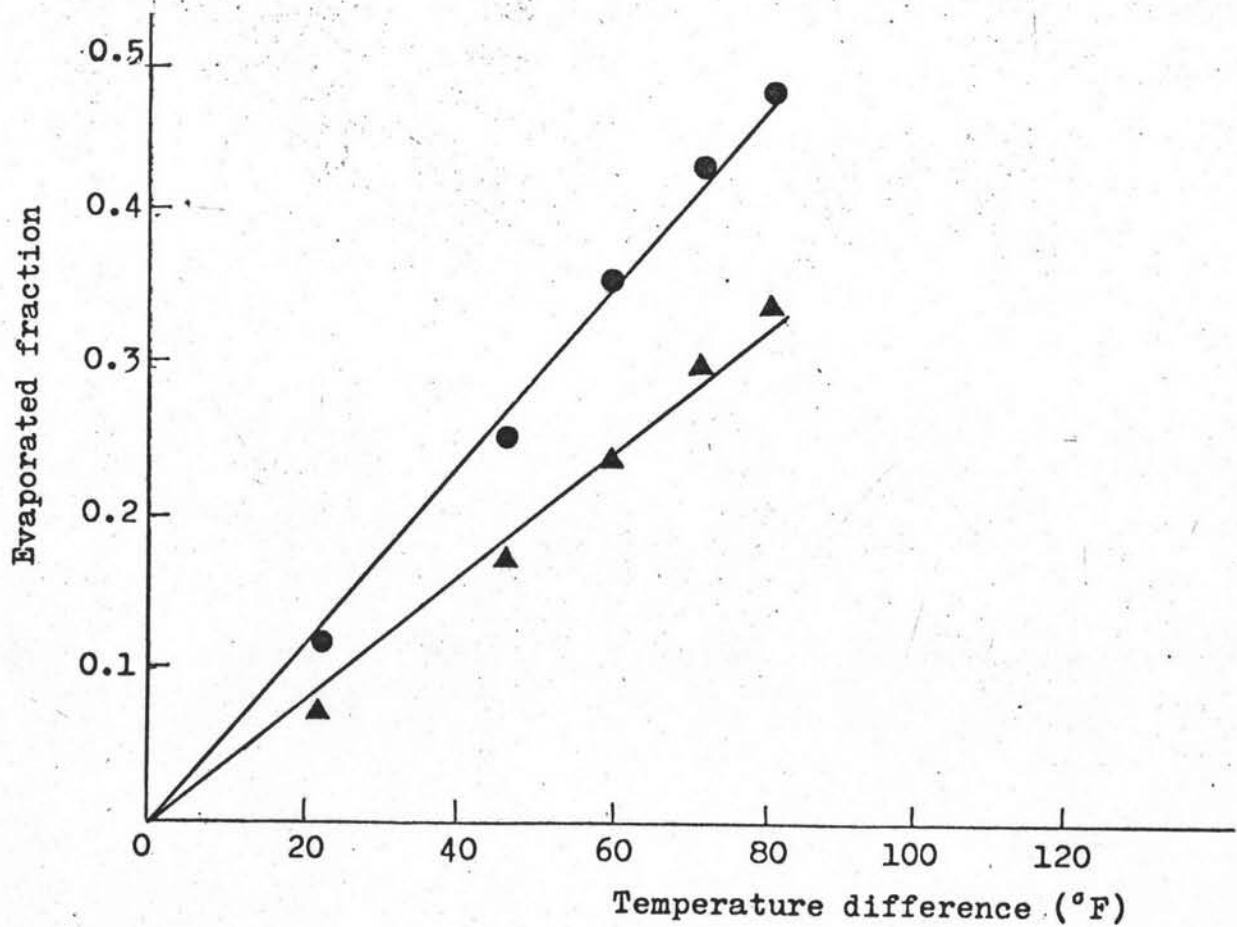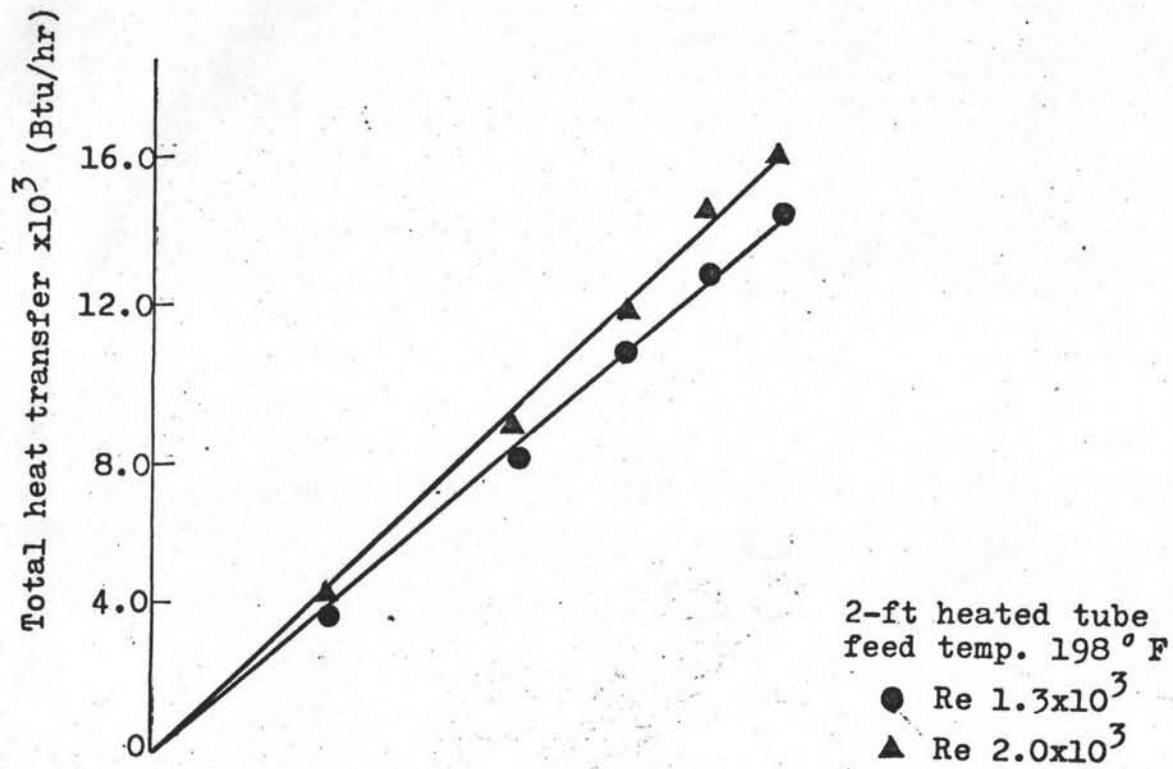
Fig.5.2 Variations of evaporated fraction and total heat
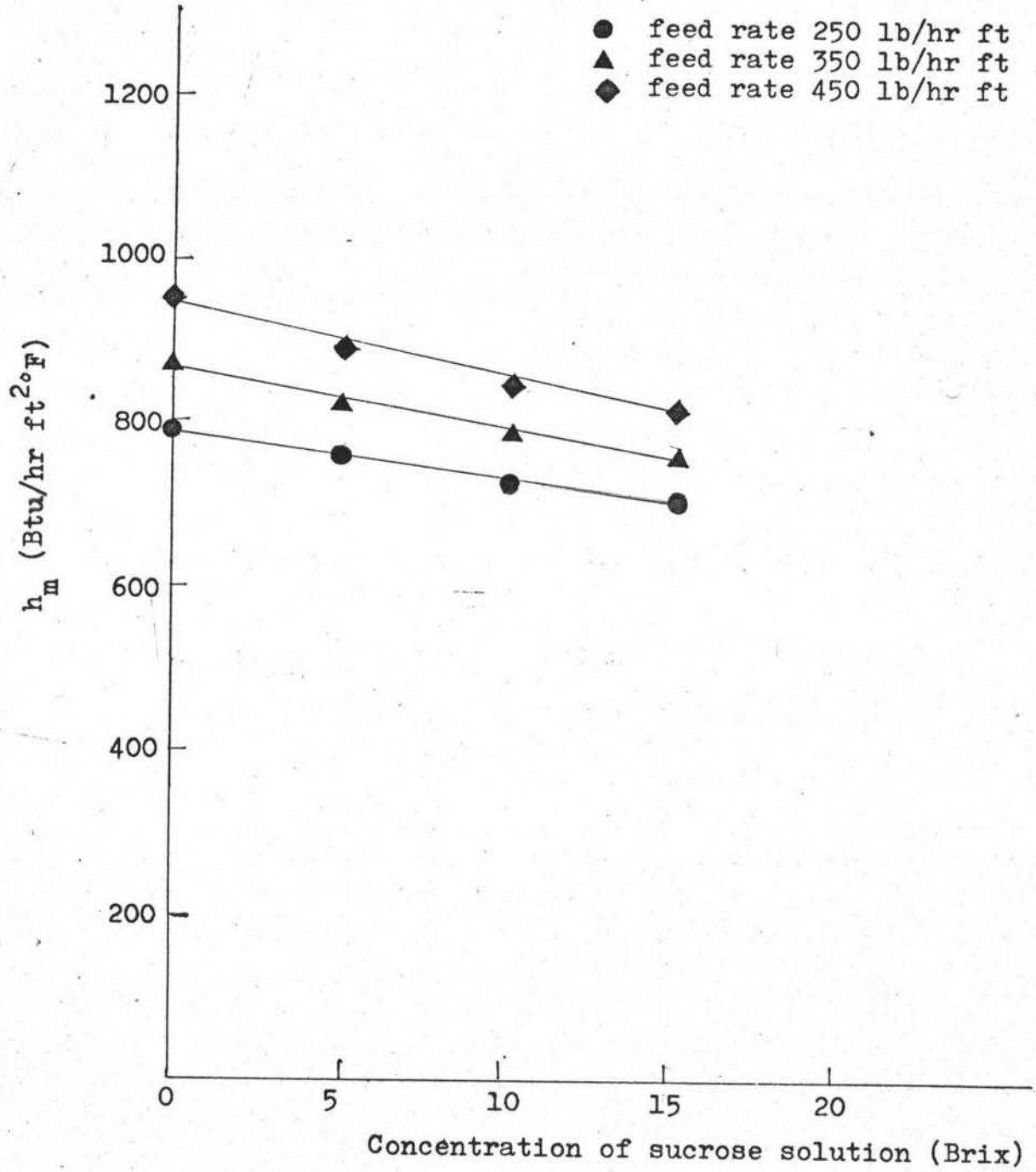transfer with temperature difference

Fig.5.3 Variation of h_m with viscosity (concentration)

concentrations of sucrose solutions is reduced into a single curve as shown in Fig 5.4. This curved may be used to predict h_m when Reynolds number is known. It is applicable for the concentrations of sucrose solution between 0-15 Brix.

The component mass balance of the evaporation of sucrose solutions is shown in Table D-8. The percentage loss is about 8%. This is due to entrainment appeared at low feed rate and high concentration of sucrose solutions; where boiling can be vigorous and small amount of solution is carried away with the vapor.

5.2.4 Effect of Feed Temperature on the Operation of the Evaporator

The effect of feed temperature on h_m was illustrated in Fig 4.6. When the feed entered the evaporating body at the temperature far from its boiling point, part of heated tube was taken up as a preheating zone. In the case of preheating, resistance of heat transfer was only the wall resistance; the heat could be transferred at a higher rate than in the case of film evaporation. Hence, heat transfer coefficient of preheating zone was higher than the heat transfer coefficient of evaporating zone. If the feed temperature is much lower than its boiling point, the preheating zone is lengthened, and h_m will be
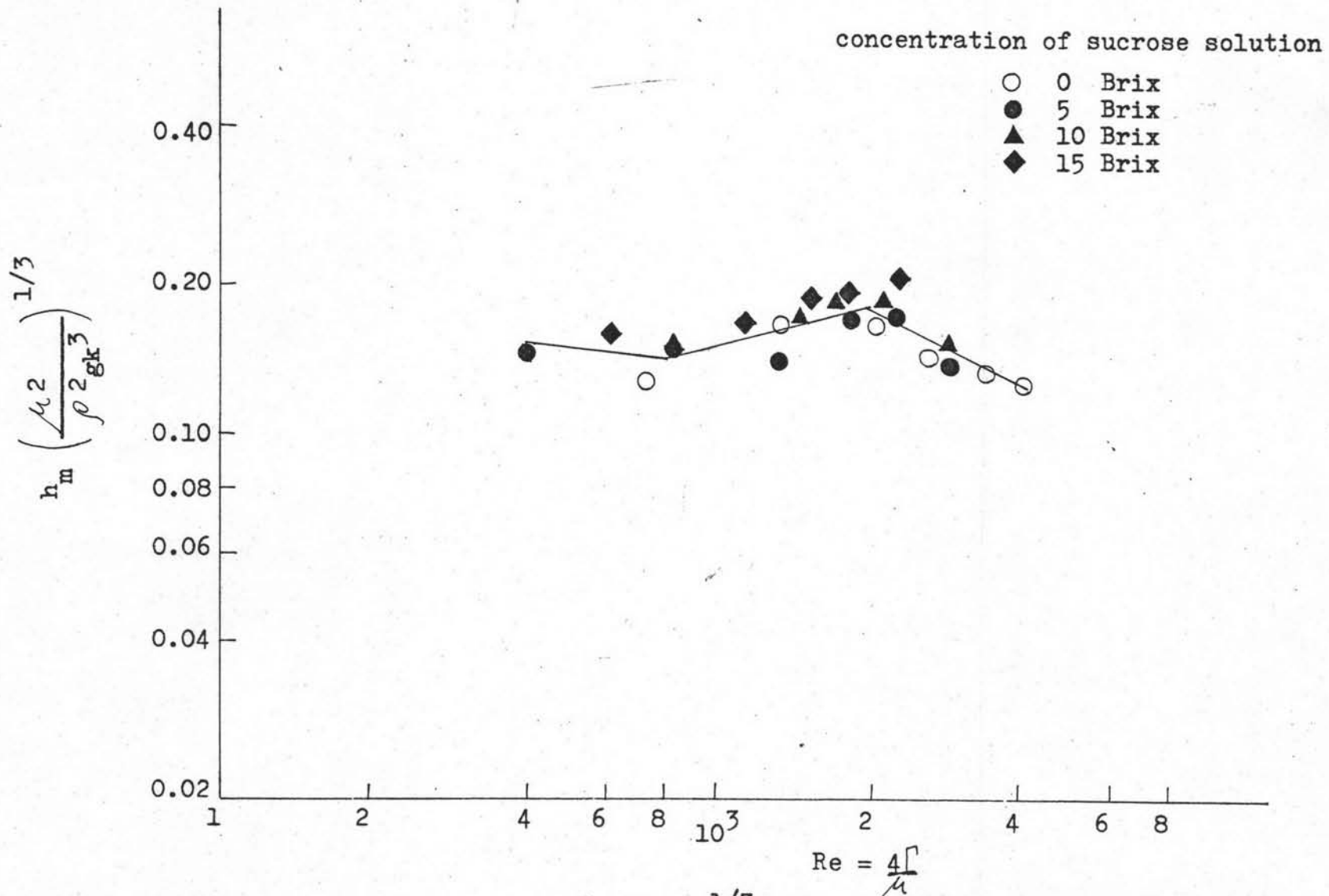
Fig.5.4 Variation of $h_m \left(\dfrac{\mu^2}{\rho^2 gk^3} \right)^{1/3}$ with Re for sucrose solution 0-15 Brix

concentration of sucrose solution

- ○ 0 Brix
- ● 5 Brix
- ▲ 10 Brix
- ◆ 15 Brix

$Re = \dfrac{4\Gamma}{\mu}$

higher. As the feed temperature is close to its boiling point heat transfer coefficient of evaporating zone dominates, and therefore, h_m is lower.

5.2.5 Effect of Length of Heated Tube on h_m

The results were ellustrated in Fig 4.7b. The mean heat transfer coefficient of a 2-ft tube appears to be higher than that of a 1-ft tube. The difference is due to the increasing of contact time. In the longer tube, contact time is greater resulting in thinner film and so higher h_m.

5.3 Performance of the Experimental Evaporator

5.3.1 Comparison with Theoretical Expression

The $\bigcap_L$ of concentrated solutions were calculated from equation 2.23, and the results were compared graphically with the experimental value in Fig 5.5. By an average method a line was fitted to the data points and the deviation of experimental values from the theoretical ones was evaluated to be about 5.7%. It can be concluded, therefore, that the theoretical expression applies well in the laminar regime.

5.3.2 Selection of Operating Condition

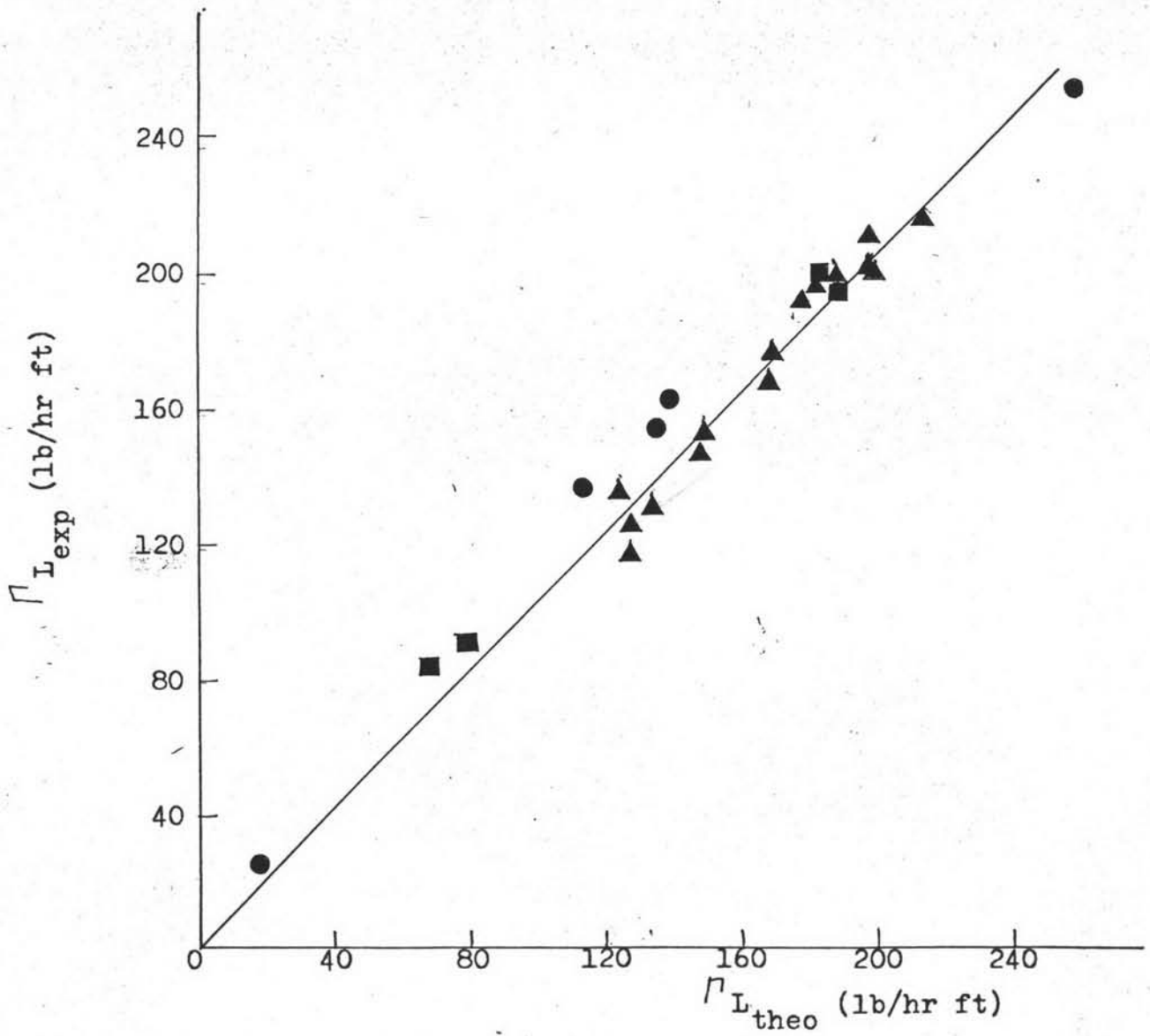It was observed from the experiments that because

Fig. 5.5 Comparison between $\Gamma_{L_{exp}}$ and $\Gamma_{L_{theo}}$

relatively long tube was employed, there was a limit to the lowest feed rate which could be used without being dried up. Evaporation of sucrose solutions at concentration of 10 and 15 Brix at a feed rate lower than 200 lb/hrft operating in a 2-ft tube and at a steam temperature of 284 ^{O}F were not carried out due to observation of apparent crystallization. The maximum feed rate was founded to be limited by "overshooting" of the feed at the distributor. Evaporation of water and sucrose solutions at a feed rate higher than 460 lb/hr ft, in a-2 ft tube, h_m dropped sharply. The suitable feed rate for the evaporator was founded to be about 460 lb/hr ft.

Effect of steam temperature on h_m was suggested the operating condition for this type of evaporator. The wall temperature must not exceed 280°F since h_m is almost constant with increasing steam temperature above that.

5.4 Viscosities of Sucrose Solutions

The measurement of viscosities of sucrose solutions were carried out at sucrose concentration of 5, 10, 15 and 20 Brix. There is little information on the variation of viscosity with temperature for such concentrations. Measurements were made and the results are shown in Fig 4.8. It was observed that viscosity varies inversely with temperature

but directly with concentration. An equation has been developed to express the relationship of viscosity; temperature and concentration in the following form:

$$\log \eta = \frac{A}{T} + B$$

It was found that viscosities of sucrose sulutions of 0-40 Brix could be represented by an equation shown below:

$$\log \eta = \frac{17.3 + 0.18c^{0.71}}{t} (0.62 - 0.04c^{1.09}) \quad (5.1)$$

when η = viscosity, centipoise

t = temperature, ^{o}C

c = concentration, Brix

The correlation together with experimental results are shown in Fig 5.6, showing good agreement.
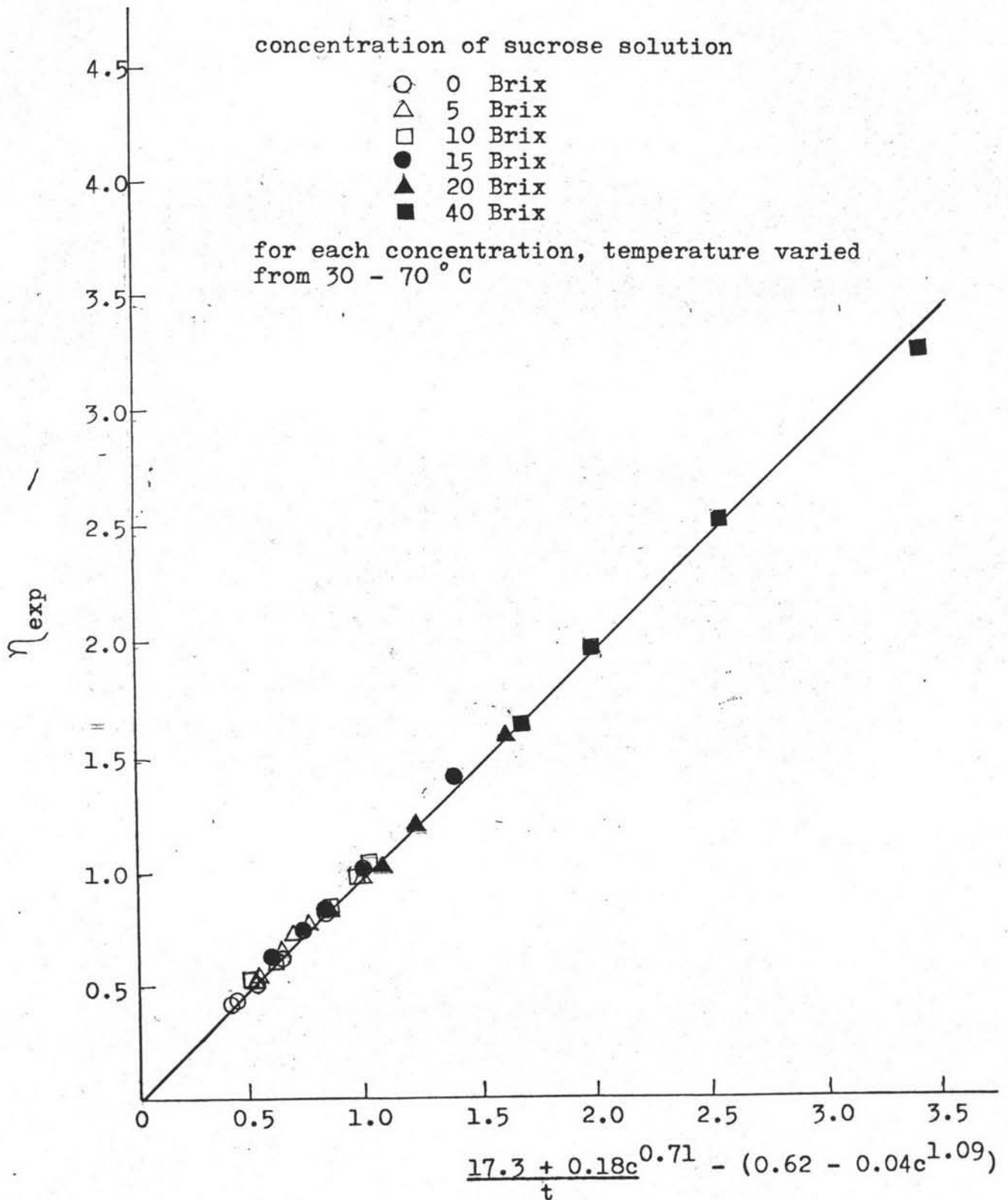
Fig.5.6 Comparison between η_{exp} and η_{eq}