4.9 Correction factor for mean temperature difference

The correction factor for the mean temperature difference is a function of two dimensionless temperature ratios:

$$R = \frac{(T_1 - T_2)}{(t_2 - t_1)} \tag{4.8}$$

R is equal to the shell-side fluid flow rate times the fluid mean specific heat, divided by the tube-side fluid flow rate times the tube-side fluid specific heat.

$$S = \frac{(t_2 - t_1)}{(T_1 - t_1)} \tag{4.9}$$

S is the measure of the temperature efficiency of the exchanger.

For 1 shell: 2tube passe exchanger, the correction factor is given by Equation 4.10 and is plotted in Figure 4.7:

$$F_t = \frac{\sqrt{(R^2+1)}\, ln\, [\frac{(1-S)}{(1-RS)}]}{(R-1)\, ln\, [\frac{2 - S[R+1-\sqrt{R^2+1}]}{2 - S[R+1+\sqrt{R^2+1}]}]} \tag{4.10}$$

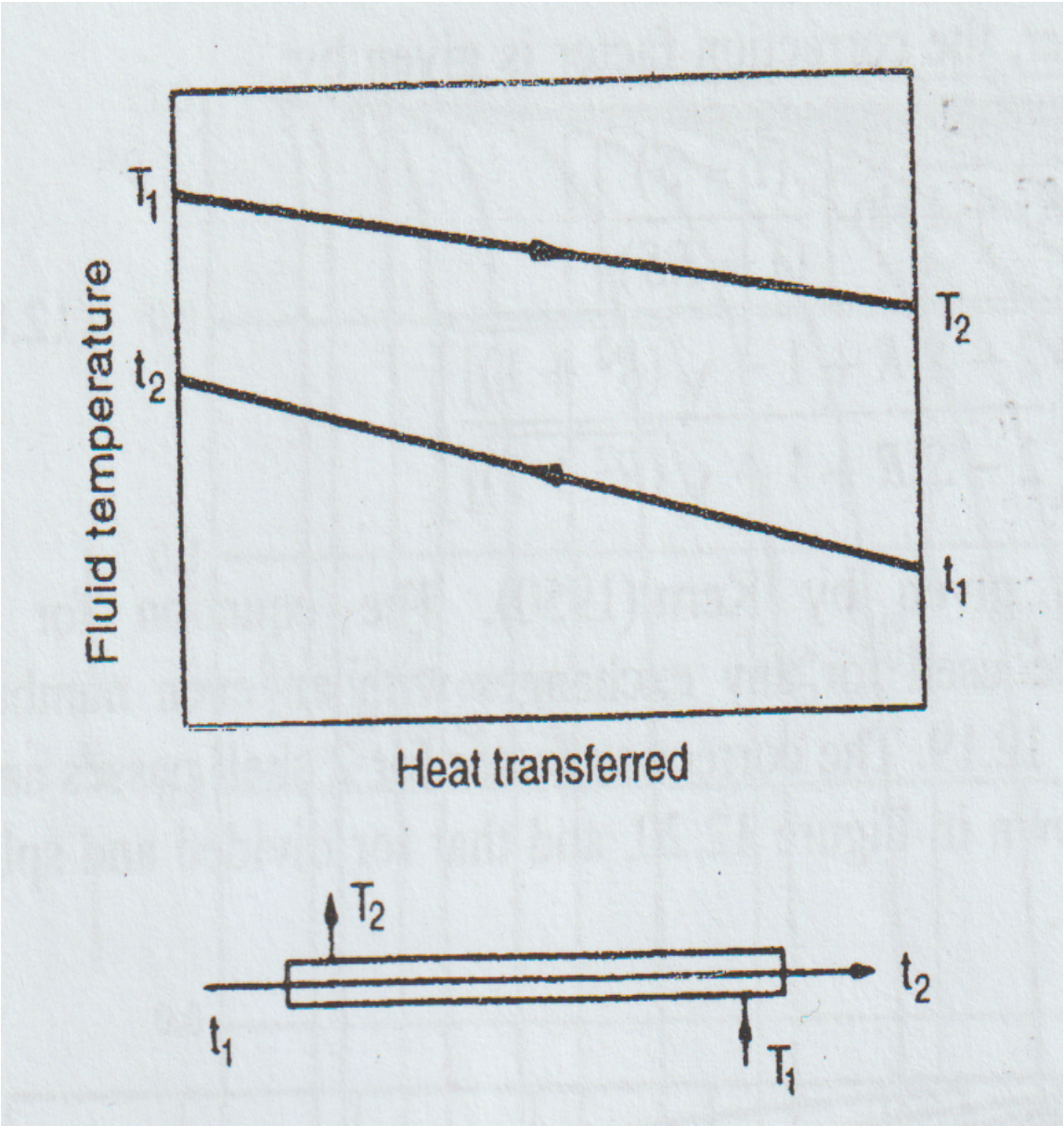

Figure 4.7: Temperature profile for counter current flow

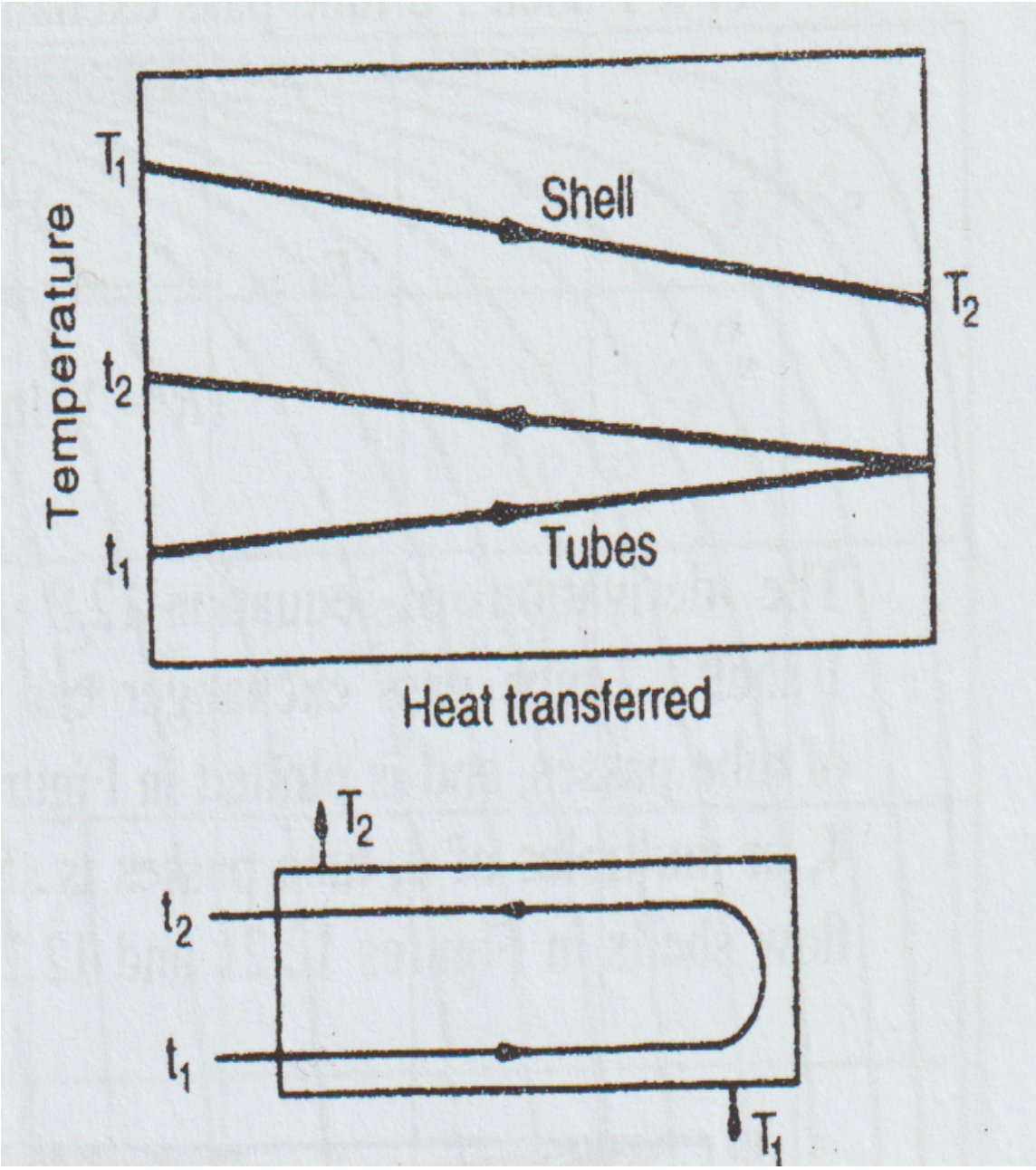

Figure 4.8: Temperature profile for 1:2 exchanger

4.10 Tube-side heat transfer coefficient and pressure drop

4.10.1 Heat transfer

4.10.1.1 Turbulent flow

Heat transfer inside conduits of uniform cross sectional area are correlated by Equation 4.11:

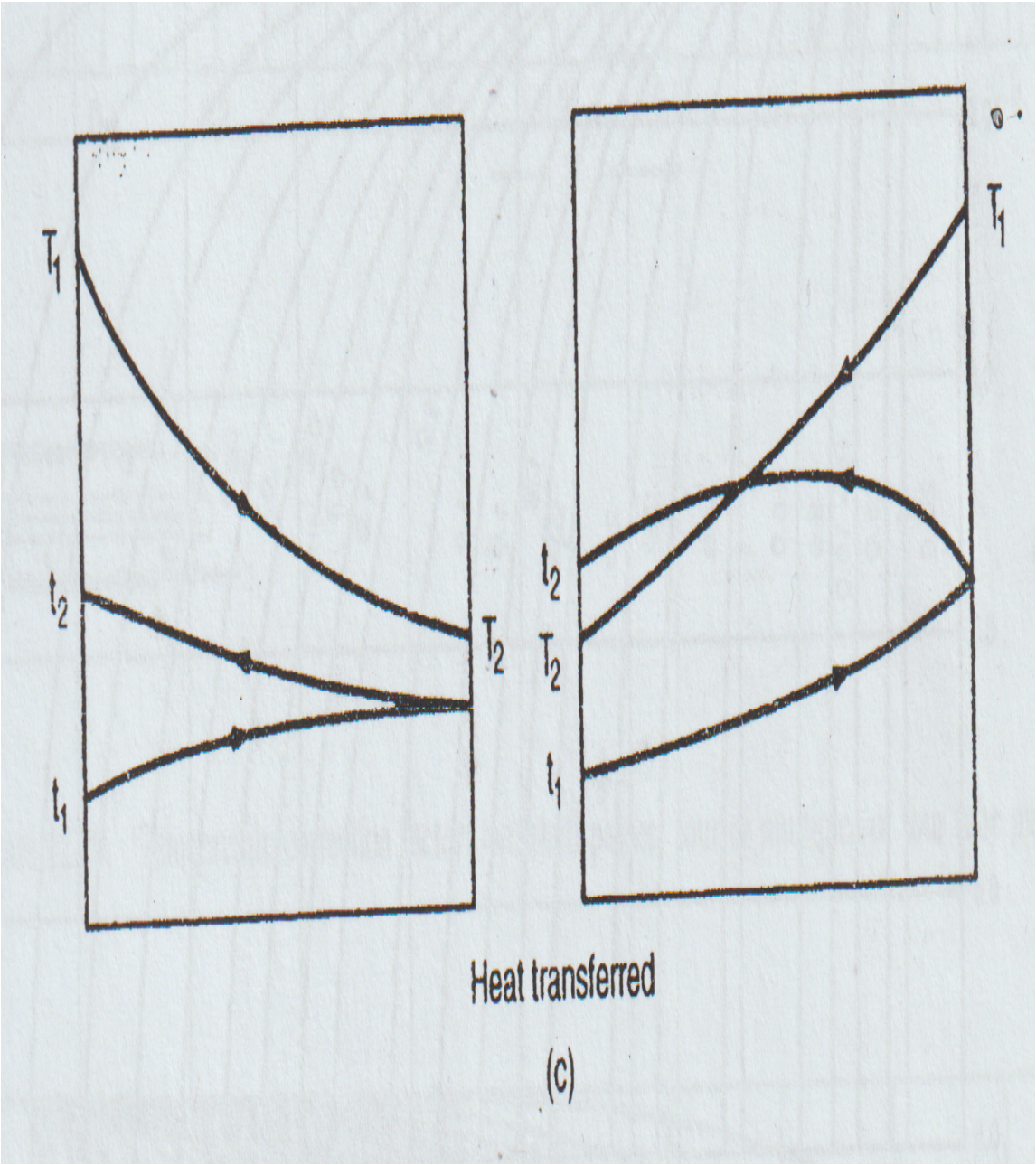

Figure 4.9: Temperature profile for cross flow

$$Nu = C\, Re^a Pr^b \left(\frac{\mu}{\mu_w}\right)^c \tag{4.11}$$

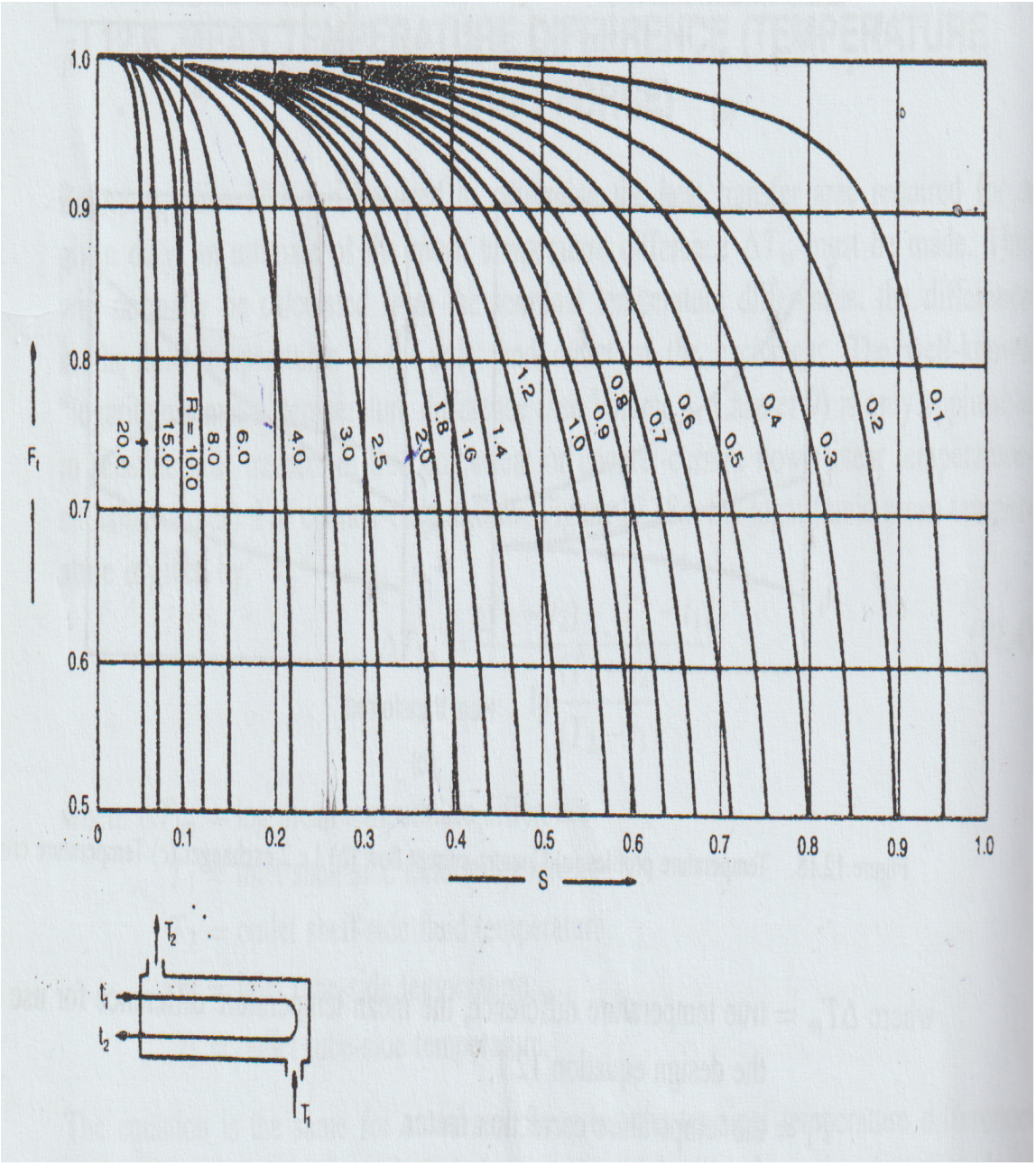

Figure 4.10: Temperature correction factor: one shell pass; two or more even tube passes

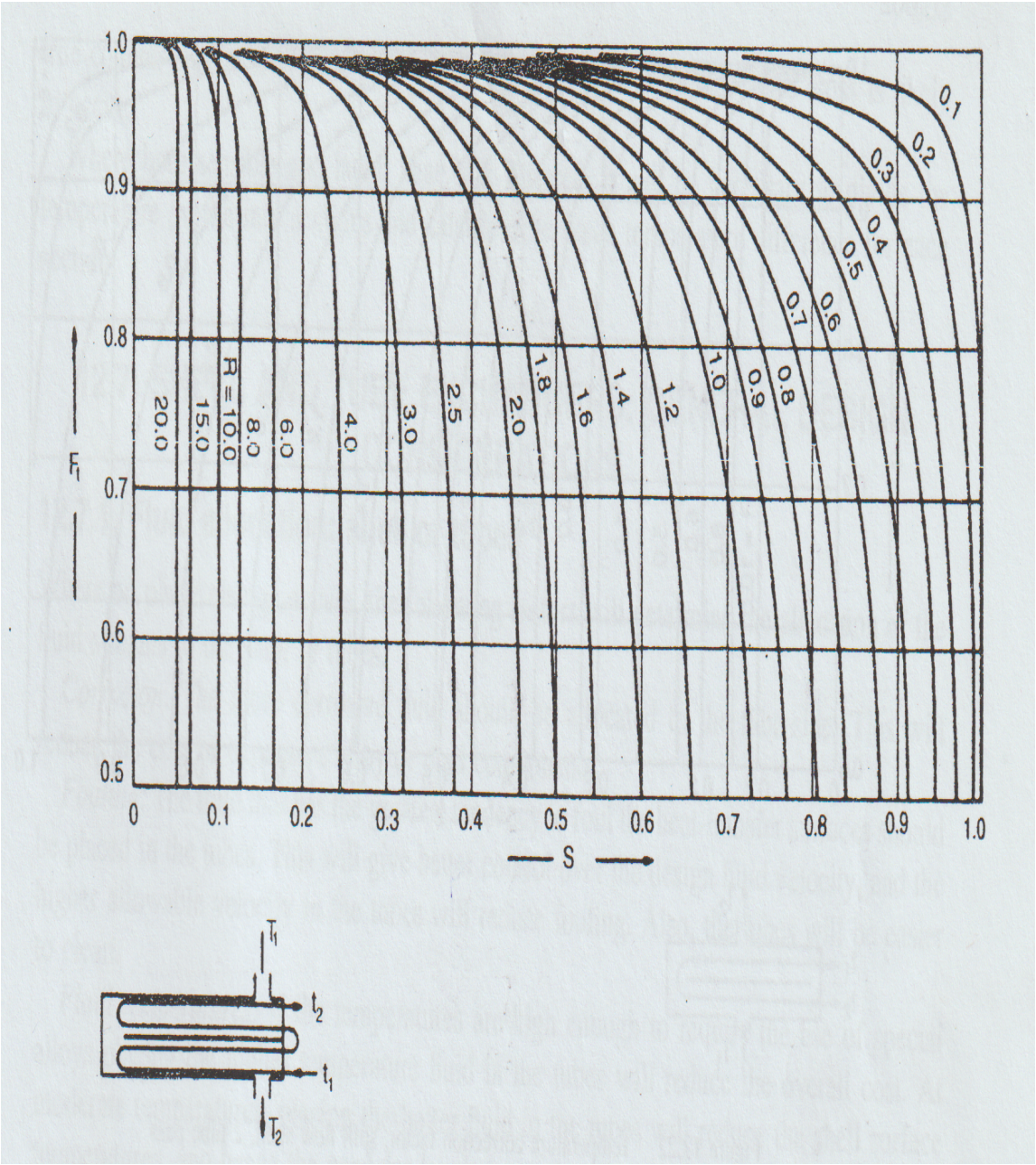

Figure 4.11: Temperature correction factor: two shell pass; four or multiples of four tube passes

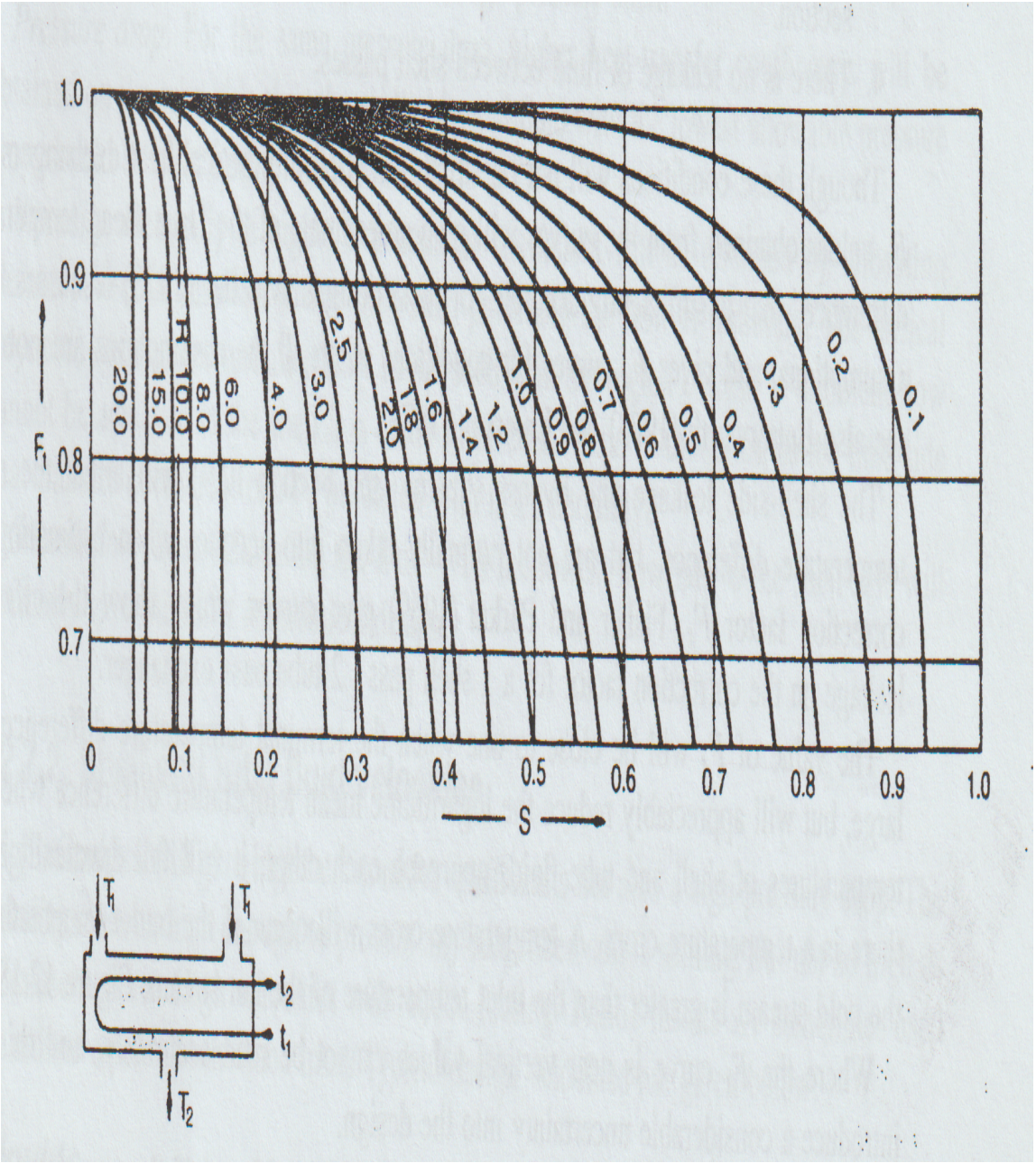

Figure 4.12: Temperature correction factor: divided-flow shell; two or more even-tube passes

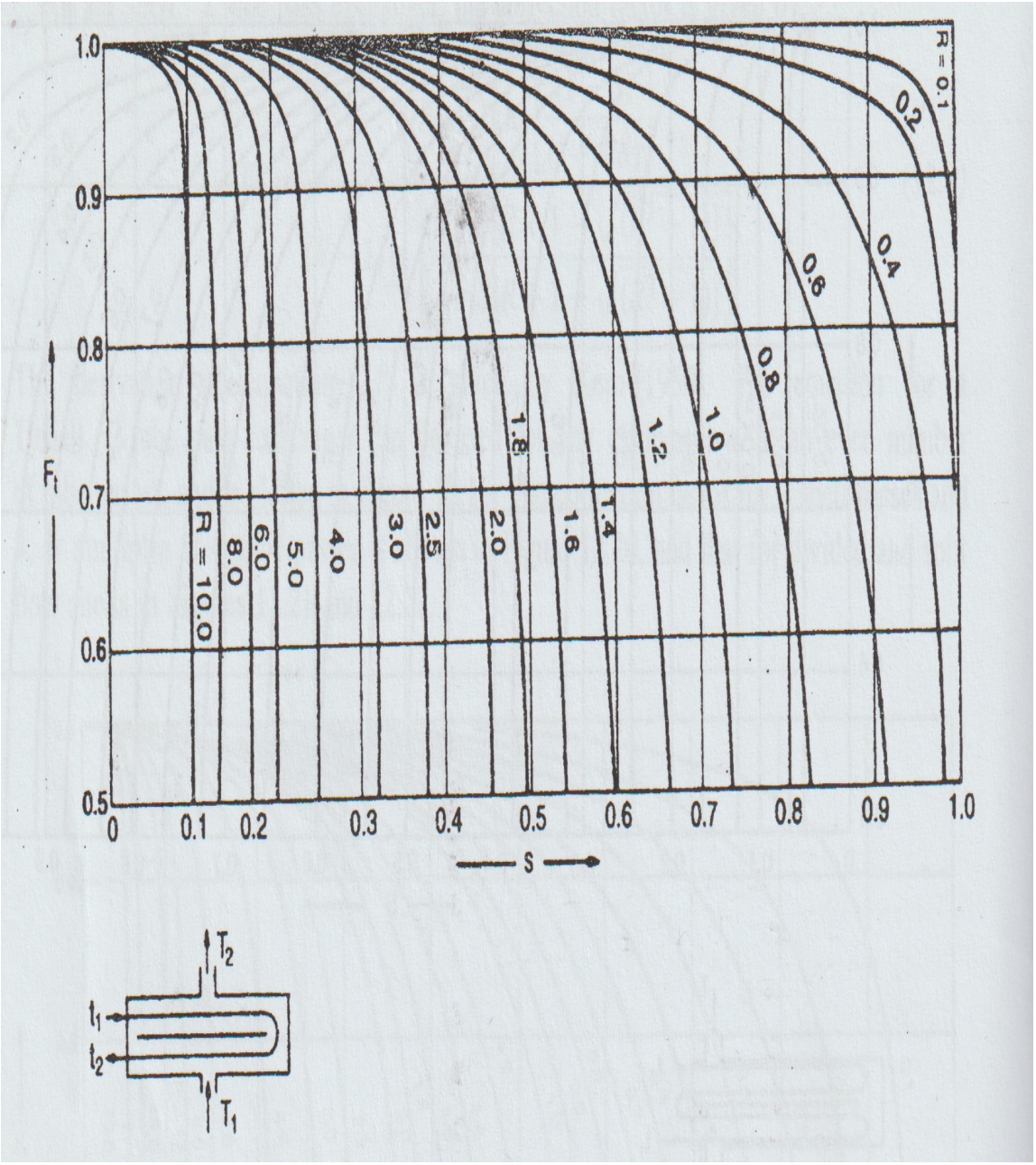

Figure 4.13: Temperature correction factor: split flow shell; 2 tube pass

where:

Nu= Nusselt number=$(h_i d_e/k_f)$

Re= Reynolds number=$(\rho u_t d_e/\mu)$=$(G_f d_e/\mu)$

Pr= Prandtl number=$(C_p \mu/k_f)$

$\mathbf{h}_i$= inside coefficient, W·m^{-2}·^{o}C^{-1}

$\mathbf{d}_e$= equivalent or hydraulic mean diameter (m)=$\dfrac{4 \times \text{cross-sectional area for flow}}{\text{Wetted perimeter}}$=d$_i$

u_t= fluid velocity (m·s^{-1})

$\mathbf{k}_f$= fluid thermal conductivity (W·m^{-1}·^{o}C^{-1})

$\mathbf{G}_t$= mass velocity, mass flow per unit area (kg·m^{-2}·s^{-1})

μ= fluid viscosity at the bulk fluid temperature (N·s·m^{-2})

μ_w= fluid viscosity at the wall

C_p= Heat capacity or fluid specific heat (J·kg^{-1}·^{o}C^{-1})

The index (a) for the Reynolds number is taken as 0.8. The index for Prandtl (b) can range from 0.3 for cooling and 0.4 for heating. For the viscosity ratio (c) is taken as 0.14.

$$Nu = C\, Re^{0.8} Pr^{0.33} (\frac{\mu}{\mu_w})^{0.14} \tag{4.12}$$

where C can be taken as 0.021 for gases, 0.023 for non-viscous liquids, and 0.027 for viscous liquids.

4.10.1.2 Laminar flow

For laminar flow (Reynolds Number<2000), Nusselt number can be estimated according to the following:

$$Nu = 1.86(RePr)^{0.33}(\frac{d_e}{L})^{0.33}(\frac{\mu}{\mu_w})^{0.14} \tag{4.13}$$

where L is the length of pipe in meters. (**Note**:) Nusselt number given by Equation 4.13 < 3.5, Nusselt number must be taken as 3.5

4.10.1.3 Transition region

Transition region should be avoided in the design of exchangers.

4.10.2 Heat-transfer factor($\mathbf{j_h}$)

The heat transfer factor is given by Equation 4.14:

$$j_h = StPr^{0.67}(\frac{\mu}{\mu_w})^{-0.14} \tag{4.14}$$

where:

St= Stanton number = (Nu/RePr)=($h_i/\rho u_t C_p$)

Equation 4.14 can be rearranged and written as follows:

$$\frac{h_i d_i}{k_f} = j_h RePr^{0.33}(\frac{\mu}{\mu_w})^{0.14} \tag{4.15}$$

4.10.3 Viscosity correction factor

This factor is significant for viscous liquids. To estimate this factor, an estimate of the wall temperature temperature is needed. The wall temperature can be estimated according to the following relation:

$$h_i(t_w - t) = U(T - t) \tag{4.16}$$

where:

t= tube-side bulk temperature (mean)

t_w= estimated wall temperature

T= shell side bulk temperature (mean)