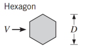
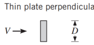
External Flow

Table 7.7 Summary of convection heat transfer correlations for external flow[a,b]

Correlation		Geometry	Conditions[c]
$\delta = 5x\,Re_x^{-1/2}$	(7.19)	Flat plate	Laminar, T_f
$C_{f,x} = 0.664\,Re_x^{-1/2}$	(7.20)	Flat plate	Laminar, local, T_f
$Nu_x = 0.332\,Re_x^{1/2}\,Pr^{1/3}$	(7.23)	Flat plate	Laminar, local, T_f, $Pr \gtrsim 0.6$
$\delta_t = \delta\,Pr^{-1/3}$	(7.24)	Flat plate	Laminar, T_f
$\overline{C}_{f,x} = 1.328\,Re_x^{-1/2}$	(7.29)	Flat plate	Laminar, average, T_f
$\overline{Nu}_x = 0.664\,Re_x^{1/2}\,Pr^{1/3}$	(7.30)	Flat plate	Laminar, average, T_f, $Pr \gtrsim 0.6$
$Nu_x = 0.564\,Pe_x^{1/2}$	(7.32)	Flat plate	Laminar, local, T_f, $Pr \lesssim 0.05$, $Pe_x \gtrsim 100$
$C_{f,x} = 0.0592\,Re_x^{-1/5}$	(7.34)	Flat plate	Turbulent, local, T_f, $Re_x \lesssim 10^8$
$\delta = 0.37x\,Re_x^{-1/5}$	(7.35)	Flat plate	Turbulent, T_f, $Re_x \lesssim 10^8$
$Nu_x = 0.0296\,Re_x^{4/5}\,Pr^{1/3}$	(7.36)	Flat plate	Turbulent, local, T_f, $Re_x \lesssim 10^8$, $0.6 \lesssim Pr \lesssim 60$
$\overline{C}_{f,L} = 0.074\,Re_L^{-1/5} - 1742\,Re_L^{-1}$	(7.40)	Flat plate	Mixed, average, T_f, $Re_{x,c} = 5 \times 10^5$, $Re_L \lesssim 10^8$
$\overline{Nu}_L = (0.037\,Re_L^{4/5} - 871)Pr^{1/3}$	(7.38)	Flat plate	Mixed, average, T_f, $Re_{x,c} = 5 \times 10^5$, $Re_L \lesssim 10^8$, $0.6 \lesssim Pr \lesssim 60$
$\overline{Nu}_D = C\,Re_D^m\,Pr^{1/3}$ (Table 7.2)	(7.52)	Cylinder	Average, T_f, $0.4 \lesssim Re_D \lesssim 4 \times 10^5$, $Pr \gtrsim 0.7$
$\overline{Nu}_D = C\,Re_D^m\,Pr^n(Pr/Pr_s)^{1/4}$ (Table 7.4)	(7.53)	Cylinder	Average, T_∞, $1 \lesssim Re_D \lesssim 10^6$, $0.7 \lesssim Pr \lesssim 500$
$\overline{Nu}_D = 0.3 + [0.62\,Re_D^{1/2}\,Pr^{1/3} \times [1 + (0.4/Pr)^{2/3}]^{-1/4}] \times [1 + (Re_D/282{,}000)^{5/8}]^{4/5}$	(7.54)	Cylinder	Average, T_f, $Re_D\,Pr \gtrsim 0.2$
$\overline{Nu}_D = 2 + (0.4\,Re_D^{1/2} + 0.06\,Re_D^{2/3})Pr^{0.4} \times (\mu/\mu_s)^{1/4}$	(7.56)	Sphere	Average, T_∞, $3.5 \lesssim Re_D \lesssim 7.6 \times 10^4$, $0.71 \lesssim Pr \lesssim 380$, $1.0 \lesssim (\mu/\mu_s) \lesssim 3.2$
$\overline{Nu}_D = 2 + 0.6\,Re_D^{1/2}\,Pr^{1/3}$	(7.57)	Falling drop	Average, T_∞
$\overline{Nu}_D = C_1C_2\,Re_{D,\max}^m\,Pr^{0.36}(Pr/Pr_s)^{1/4}$ (Tables 7.5, 7.6)	(7.58), (7.59)	Tube bank[d]	Average, $\overline{T}$, $10 \lesssim Re_D \lesssim 2 \times 10^6$, $0.7 \lesssim Pr \lesssim 500$

[d]For tube banks and packed beds, properties are evaluated at the average fluid temperature, $\overline{T} = (T_i + T_o)/2$.

Table 7.3 Constants of Equation 7.52 for noncircular cylinders in cross flow of a gas [13, 14][a]

Geometry		Re_D	C	m
Square 		6000–60,000	0.304	0.59
		5000–60,000	0.158	0.66
Hexagon 		5200–20,400	0.164	0.638
		20,400–105,000	0.039	0.78
		4500–90,700	0.150	0.638
Thin plate perpendicular to flow 	Front	10,000–50,000	0.667	0.500
	Back	7000–80,000	0.191	0.667

[a]These tabular values are based on the recommendations of Sparrow et al. [14] for air, with extension to other fluids through the $Pr^{1/3}$ dependence of Equation 7.52. A Prandtl number of $Pr = 0.7$ was assumed for the experimental results for air that are described in [14].

Table 7.2 Constants of Equation 7.52 for the circular cylinder in cross flow [11, 12]

Re_D	C	m
0.4–4	0.989	0.330
4–40	0.911	0.385
40–4000	0.683	0.466
4000–40,000	0.193	0.618
40,000–400,000	0.027	0.805

Table 7.4 Constants of Equation 7.53 for the circular cylinder in cross flow [17]

Re_D	C	m
1–40	0.75	0.4
40–1000	0.51	0.5
10^3–2×10^5	0.26	0.6
2×10^5–10^6	0.076	0.7

计算流程

$$Re_{L长度} = \frac{\rho VL}{\mu} = \frac{VL}{\nu}$$

$$Re_{D直径} = \frac{\rho VD}{\mu} = \frac{VD}{\nu}$$

$$\bar{h} = \frac{k}{L} \cdot \overline{Nu}$$

$$q = h \cdot A_s \cdot (T_s - T_\infty)$$

Internal Flow (Tube)

Table 8.4 Summary of convection correlations for flow in a circular tube[a,b,e]

Correlation		Conditions
$f = 64/Re_D$	(8.19)	Laminar, fully developed
$Nu_D = 4.36$	(8.53)	Laminar, fully developed, uniform q_s'' 恒定热通量
$Nu_D = 3.66$	(8.55)	Laminar, fully developed, uniform T_s 恒定表面温度
$\overline{Nu}_D = 3.66 + \dfrac{0.0668\,Gz_D}{1 + 0.04\,Gz_D^{2/3}}$	(8.57)	Laminar, thermal entry (or combined entry with $Pr \gtrsim 5$), uniform T_s, $Gz_D = (D/x)\,Re_D\,Pr$
$\overline{Nu}_D = \dfrac{\dfrac{3.66}{\tanh[2.264\,Gz_D^{-1/3} + 1.7\,Gz_D^{-2/3}]} + 0.0499\,Gz_D\tanh(Gz_D^{-1})}{\tanh(2.432\,Pr^{1/6}\,Gz_D^{-1/6})}$	(8.58)	Laminar, combined entry, $Pr \gtrsim 0.1$, uniform T_s, $Gz_D = (D/x)\,Re_D\,Pr$
$Nu_D = 0.023\,Re_D^{4/5}\,Pr^n$ 湍流混合比较均匀，所以Nu局部=平均	(8.60)[d]	Turbulent, fully developed, $0.6 \lesssim Pr \lesssim 160$, $Re_D \gtrsim 10{,}000$, $(L/D) \gtrsim 10$, $n = 0.4$ for $T_s > T_m$ and $n = 0.3$ for $T_s < T_m$
$Nu_D = 0.027\,Re_D^{4/5}\,Pr^{1/3}\left(\dfrac{\mu}{\mu_s}\right)^{0.14}$	(8.61)[d]	Turbulent, fully developed, $0.7 \lesssim Pr \lesssim 16{,}700$, $Re_D \gtrsim 10{,}000$, $L/D \gtrsim 10$
$Nu_D = \dfrac{(f/8)(Re_D - 1000)\,Pr}{1 + 12.7(f/8)^{1/2}(Pr^{2/3} - 1)}$	(8.62)[d]	Turbulent, fully developed, $0.5 \lesssim Pr \lesssim 2000$, $3000 \lesssim Re_D \lesssim 5 \times 10^6$, $(L/D) \gtrsim 10$
$Nu_D = 4.82 + 0.0185(Re_D\,Pr)^{0.827}$	(8.64)	Liquid metals, turbulent, fully developed, uniform q_s'', $3.6 \times 10^3 \lesssim Re_D \lesssim 9.05 \times 10^5$, $3 \times 10^{-3} \lesssim Pr \lesssim 5 \times 10^{-2}$, $10^2 \lesssim Re_D\,Pr \lesssim 10^4$
$Nu_D = 5.0 + 0.025(Re_D\,Pr)^{0.8}$	(8.65)	Liquid metals, turbulent, fully developed, uniform T_s, $Re_D\,Pr \gtrsim 100$
$\dfrac{1}{\sqrt{f}} = -2.0\log\left[\dfrac{e/D}{3.7} + \dfrac{2.51}{Re_D\sqrt{f}}\right]$	(8.20)[c]	Turbulent, fully developed
$f = (0.790\ln Re_D - 1.64)^{-2}$	(8.21)[c]	Turbulent, fully developed, smooth walls, $3000 \lesssim Re_D \lesssim 5 \times 10^6$

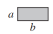
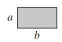
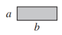
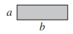
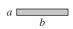
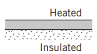
[a]Properties in Equations 8.53, 8.55, 8.60, 8.61, 8.62, 8.64, and 8.65 are based on T_m; properties in Equations 8.19, 8.20, and 8.21 are based on $T_f = (T_s + T_m)/2$; properties in Equations 8.57 and 8.58 are based on $\overline{T}_m = (T_{m,i} + T_{m,o})/2$.

[c]Equation 8.20 pertains to smooth or rough tubes. Equation 8.21 pertains to smooth tubes.

[d]As a first approximation, Equations 8.60, 8.61, or 8.62 may be used to evaluate the average Nusselt number $\overline{Nu}_D$ over the entire tube length, if $(L/D) \gtrsim 10$. The properties should then be evaluated at the average of the mean temperature, $\overline{T}_m = (T_{m,i} + T_{m,o})/2$.

[e]For tubes of noncircular cross section, $Re_D \equiv D_h u_m/\nu$, $D_h \equiv 4A_c/P$, and $u_m = \dot{m}/\rho A_c$. Results for fully developed laminar flow are provided in Table 8.1. For turbulent flow, Equation 8.60 may be used as a first approximation.

Table 8.1 Nusselt numbers and friction factors for fully developed laminar flow in tubes of differing cross section

Cross Section	$\dfrac{b}{a}$	$Nu_D \equiv \dfrac{hD_h}{k}$ (Uniform q_s'')	$Nu_D \equiv \dfrac{hD_h}{k}$ (Uniform T_s)	$f\,Re_{D_h}$
	—	4.36	3.66	64
	1.0	3.61	2.98	57
	1.43	3.73	3.08	59
	2.0	4.12	3.39	62
	3.0	4.79	3.96	69
	4.0	5.33	4.44	73
	8.0	6.49	5.60	82
	∞	8.23	7.54	96
Heated  Insulated	∞	5.39	4.86	96
	—	3.11	2.49	53

$$D_h = \frac{4A_c\ 截面积}{P\ 截面周长}$$

Hydraulic diameter

截面不是圆形的近似

NTU Method

Flow Arrangement	Relation		
Parallel flw	$\varepsilon = \dfrac{1 - \exp[-\text{NTU}(1 + C_r)]}{1 + C_r}$		(11.28a)
Counterflw	$\varepsilon = \dfrac{1 - \exp[-\text{NTU}(1 - C_r)]}{1 - C_r\exp[-\text{NTU}(1 - C_r)]}$	$(C_r < 1)$	
	$\varepsilon = \dfrac{\text{NTU}}{1 + \text{NTU}}$	$(C_r = 1)$	(11.29a)
Shell-and-tube			
One shell pass (2, 4, . . . tube passes)	$\varepsilon_1 = 2\left\{1 + C_r + (1 + C_r^2)^{1/2} \times \dfrac{1 + \exp[-(\text{NTU})_1(1 + C_r^2)^{1/2}]}{1 - \exp[-(\text{NTU})_1(1 + C_r^2)^{1/2}]}\right\}^{-1}$		(11.30a)
n shell passes ($2n$, $4n$, . . . tube passes)	$\varepsilon = \left[\left(\dfrac{1 - \varepsilon_1 C_r}{1 - \varepsilon_1}\right)^n - 1\right]\left[\left(\dfrac{1 - \varepsilon_1 C_r}{1 - \varepsilon_1}\right)^n - C_r\right]^{-1}$		(11.31a)
Cross-flw (single pass)			
Both fluids unmixed	$\varepsilon = 1 - \exp\left[\left(\dfrac{1}{C_r}\right)(\text{NTU})^{0.22}\{\exp[-C_r(\text{NTU})^{0.78}] - 1\}\right]$		(11.32)
$C_{\max}$ (mixed), $C_{\min}$ (unmixed)	$\varepsilon = \left(\dfrac{1}{C_r}\right)(1 - \exp\{-C_r[1 - \exp(-\text{NTU})]\})$		(11.33a)
$C_{\min}$ (mixed), $C_{\max}$ (unmixed)	$\varepsilon = 1 - \exp(-C_r^{-1}\{1 - \exp[-C_r(\text{NTU})]\})$		(11.34a)
All exchangers ($C_r = 0$)	$\varepsilon = 1 - \exp(-\text{NTU})$		(11.35a)
Parallel flw	$\text{NTU} = -\dfrac{\ln[1 - \varepsilon(1 + C_r)]}{1 + C_r}$		(11.28b)
Counterflw	$\text{NTU} = \dfrac{1}{C_r - 1}\ln\left(\dfrac{\varepsilon - 1}{\varepsilon C_r - 1}\right)$	$(C_r < 1)$	
	$\text{NTU} = \dfrac{\varepsilon}{1 - \varepsilon}$	$(C_r = 1)$	(11.29b)
Shell-and-tube			
One shell pass (2, 4, . . . tube passes)	$(\text{NTU})_1 = -(1 + C_r^2)^{-1/2}\ln\left(\dfrac{E - 1}{E + 1}\right)$		(11.30b)
	$E = \dfrac{2/\varepsilon_1 - (1 + C_r)}{(1 + C_r^2)^{1/2}}$		(11.30c)
n shell passes ($2n$, $4n$, . . . tube passes)	Use Equations 11.30b and 11.30c with $\varepsilon_1 = \dfrac{F - 1}{F - C_r}$ $F = \left(\dfrac{\varepsilon C_r - 1}{\varepsilon - 1}\right)^{1/n}$ $\text{NTU} = n(\text{NTU})_1$		(11.31b, c, d)
Cross-flw (single pass)			
$C_{\max}$ (mixed), $C_{\min}$ (unmixed)	$\text{NTU} = -\ln\left[1 + \left(\dfrac{1}{C_r}\right)\ln(1 - \varepsilon C_r)\right]$		(11.33b)
$C_{\min}$ (mixed), $C_{\max}$ (unmixed)	$\text{NTU} = -\left(\dfrac{1}{C_r}\right)\ln[C_r\ln(1 - \varepsilon) + 1]$		(11.34b)
All exchangers ($C_r = 0$)	$\text{NTU} = -\ln(1 - \varepsilon)$		(11.35b)

$$\text{NTU} = \frac{UA}{C_{\min}}$$

$$\text{NTU} = f(\varepsilon, C_r)$$

$$\varepsilon = \frac{q}{q_{Max}} \qquad C_r = \frac{C_{Min}}{C_{Max}}$$

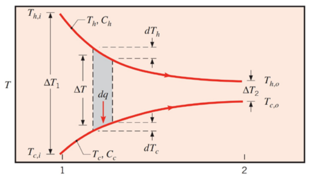
$$q_{Max} = C_{Min}\Delta T_{Max}$$

LMTD Method

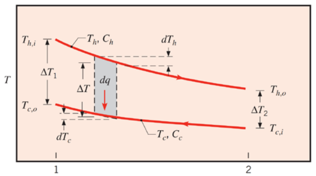
$$q = UA\Delta T_{lm} \qquad q_h = \dot{m}_h c_{p,h}\Delta T_h \qquad q_c = \dot{m}_c c_{p,c}\Delta T_c$$

$$\Delta T_{lm} = \frac{\Delta T_1 - \Delta T_2}{ln(\Delta T_1/\Delta T_2)}$$

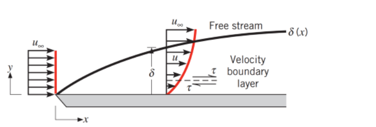
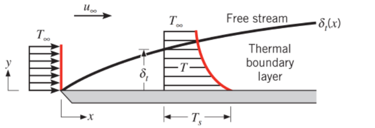
Parallel Flow



Counter Flow



Boundary Layer

速度边界层	温度边界层
	
$\delta_{边界层厚度} = \frac{0-u(y)}{0-u_\infty} = 0.99$	$\delta_{边界层厚度} = \frac{T_s - T(y)}{T_s - T_\infty} = 0.99$

$$\tau_s = \mu \left.\frac{\partial u}{\partial y}\right|_{y=0}$$

修改雷诺相似 Modified Reynold Analogy

$$F_D = \int_{A_s} \tau_s dA_s$$

扩展到 Pr 不需要 = 1

使用条件

$$C_f = \frac{\tau_s}{\frac{1}{2}\rho u_\infty}$$

层流	湍流
$\frac{dp^*}{dx^*} \approx 0$	$\frac{dp^*}{dx^*}$ 无限制

使用公式

$$\frac{C_f}{2} = St \cdot Pr^{\frac{2}{3}} \quad 0.6 < Pr < 60$$

局部 对流传热系数 h，因为 $\frac{\partial T}{\partial v}$ 在变化

conduction

$$\left.\begin{aligned} q_s'' &= -k_{流体} \left.\frac{\partial T}{\partial y}\right|_{y=0} \\ q_s'' &= h(T_s - T_\infty) \end{aligned}\right\} \Rightarrow h_{局部} = \frac{-k_{流体} \left.\frac{\partial T}{\partial y}\right|_{y=0}}{T_s - T_\infty}$$

convection

平均 对流传热系数

$$\bar{h}_{平均} = \frac{1}{A_s}\int_{A_s} h_{局部} dA_s$$

Free Convection

竖直平面

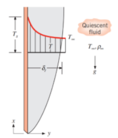
Laminar Flow

$$\overline{Nu}_L = 0.68 + \frac{0.67\,Ra_L^{\frac{1}{4}}}{\left[1+\left(\frac{0.492}{Pr}\right)^{\frac{9}{16}}\right]^{\frac{4}{9}}}$$

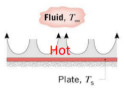
All Condition

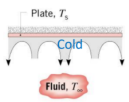
$$\overline{Nu}_L = \left\{0.825 + \frac{0.387\,Ra_L^{\frac{1}{6}}}{\left[1+\left(\frac{0.492}{Pr}\right)^{\frac{9}{16}}\right]^{\frac{8}{27}}}\right\}^2$$

All these properties are evaluated at film temperature



水平平面



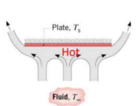


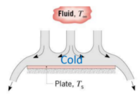
$T_s > T_\infty$ $\qquad$ $T_s < T_\infty$

$$\overline{Nu}_L = 0.54\,Ra_L^{1/4} \quad \left(10^4 < Ra_L < 10^7\right)$$

$$\overline{Nu}_L = 0.15\,Ra_L^{1/3} \quad \left(10^7 < Ra_L < 10^{11};\ \text{all } Pr\right)$$

where $L = A_s/P$.





$T_s > T_\infty$ $\qquad$ $T_s < T_\infty$

$$\overline{Nu}_L = 0.52\,Ra_L^{1/5} \quad \left(10^4 < Ra_L < 10^9;\ Pr > 0.7\right)$$

Mixed Convection

- Heat Transfer Correlations for Mixed Convection:

$$Nu^n \approx Nu_{FC}^n \pm Nu_{NC}^n$$

Nu_{FC} → Nusselt number for forced convection
Nu_{NC} → Nusselt number for natural (free) convection

+ → assisting and transverse flows
- → opposing flows

$n \approx 3$ normally but there are more values of possible n depending on the configurations

Internal Flow (Tube)

速度边界层 Hydrodynamic	温度边界层 Thermal
完全发展流 → 入口长度 Fully developed Flow → Entry Length	**完全发展流 → 入口长度** Fully developed Flow → Entry Length
$\left(\frac{x_{fd,h}}{D}\right)_{层流} \approx 0.05\,Re_D$	$\left(\frac{x_{fd,t}}{D}\right)_{层流} \approx 0.05 Re_D Pr$
$10 \lesssim \left(\frac{x_{fd,h}}{D}\right)_{湍流} \lesssim 60$	$10 \lesssim \left(\frac{x_{fd,t}}{D}\right)_{湍流}$

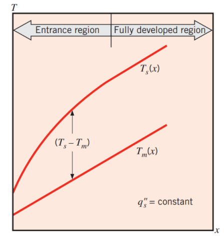
管道表面 恒定热通量 q_s'' = 定值

$$T_m(x) = T_{m,i} + \frac{q_s'' P}{\dot{m} c_p} x$$

P = 截面周长

q''_s = 管道表面热通量

$$q_{对流,总} = q_s'' \cdot PL$$



管道表面 恒定温度 T_s = 定值

Internal Flow

$$\frac{T_s - T_m(x)}{T_s - T_{m,i}} = \exp\left(-\frac{Px}{\dot{m}c_p}\bar{h}\right)$$

"External" Flow

$$\frac{\Delta T_{出}}{\Delta T_{进}} = \frac{T_\infty - T_{平均,出}}{T_\infty - T_{平均,进}} = \exp\left(-\frac{\bar{U}A_s}{\dot{m}c_p}\right)$$

$$\Delta T_{lm} = \frac{\Delta T_{出} - \Delta T_{进}}{\ln(\Delta T_{出}/\Delta T_{进})}$$

$$q_{conv} = \bar{h} A_s \Delta T_{lm} \qquad q = \bar{U} A_s \Delta T_{lm}$$

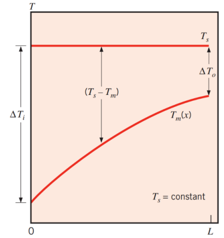
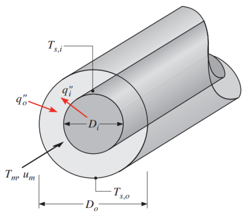


TABLE 8.2 Nusselt number for fully developed laminar flow in a circular tube annulus with one surface insulated and the other at constant temperature

D_i/D_o	Nu_i	Nu_o	Comments
0	—	3.66	See Equation 8.55
0.05	17.46	4.06	
0.10	11.56	4.11	
0.25	7.37	4.23	
0.50	5.74	4.43	
≈1.00	4.86	4.86	See Table 8.1, $b/a \to \infty$

TABLE 8.3 Influence coefficients for fully developed laminar flow in a circular tube annulus with uniform heat flux maintained at both surfaces

D_i/D_o	Nu_{ii}	Nu_{oo}	θ_i^*	θ_o^*
0	—	4.364[a]	∞	0
0.05	17.81	4.792	2.18	0.0294
0.10	11.91	4.834	1.383	0.0562
0.20	8.499	4.833	0.905	0.1041
0.40	6.583	4.979	0.603	0.1823
0.60	5.912	5.099	0.473	0.2455
0.80	5.58	5.24	0.401	0.299
1.00	5.385	5.385[b]	0.346	0.346



Case 1 普通热交换器

Conduction 热传递

$$\frac{1}{UA} = \frac{1}{(hA)_{冷端}} + R_{管壁} + \frac{1}{(hA)_{热端}}$$

Convection 热对流

Case 2 污染

$$\frac{1}{UA} = \frac{1}{(hA)_{冷端}} + \frac{R''_{f,冷端}}{A_{冷端}} + R_{管壁} + \frac{R''_{f,热端}}{A_{热端}} + \frac{1}{(hA)_{热端}}$$

Case 3 污染+鳍片

$$\frac{1}{UA} = \frac{1}{(\eta_o hA)_{冷端}} + \frac{R''_{f,冷端}}{(\eta_o A)_{冷端}} + R_{管壁} + \frac{R''_{f,热端}}{(\eta_o A)_{热端}} + \frac{1}{(\eta_o hA)_{热端}}$$

$$\eta_o = 1 - \frac{A_{fin}}{A}(1-\eta_{fin}) \qquad R_{管壁} = \frac{\ln(D_o/D_i)}{2\pi Lk}$$

Overall surface efficiency of fin array

$A = A_t$ = 总表面积 (fin + base)

A_{fin} = 鳍片面积

$$\eta_{fin} = \frac{\tanh mL}{mL} \qquad m = \sqrt{\frac{2h}{kt}}$$

Case 4 污染+鳍片 (整合版)

$$\frac{1}{UA} = \frac{1}{(\eta_o U_p A)_{冷端}} + R_{管壁} + \frac{1}{(\eta_o U_p A)_{热端}}$$

$$U_p = \frac{h}{1 + hR_f''}$$ Partial overall coefficient 部分整合系数

Dimensionless Number

St = Stanton number 斯坦顿数

$$St = \frac{h}{\rho V c_p} = \frac{Nu}{Re \cdot Pr}$$

Gr = Grashof number 格拉晓夫数

$$Gr_L = Re_L^2$$

$$Gr_L = \frac{g\beta(T_s - T_\infty)L^3}{\nu^2}$$

Pr = Prandtl number 普朗特数

$$Pr = \frac{c_p\mu}{k} = \frac{\nu_{粘度}}{\alpha_{热扩散率}}$$

Ra = Rayleigh number 瑞利数

$$Ra_L = Gr_L \cdot Pr = \left(\frac{UL}{\nu}\right)^2\left(\frac{\nu}{\alpha}\right) = \frac{U^2L^2}{\nu\alpha} = \frac{g\beta(T_s - T_\infty)L^3}{\nu\alpha}$$

α = 热扩散率

$$\beta = \frac{1}{T_f}$$