CH402 Chemical Engineering Process Design

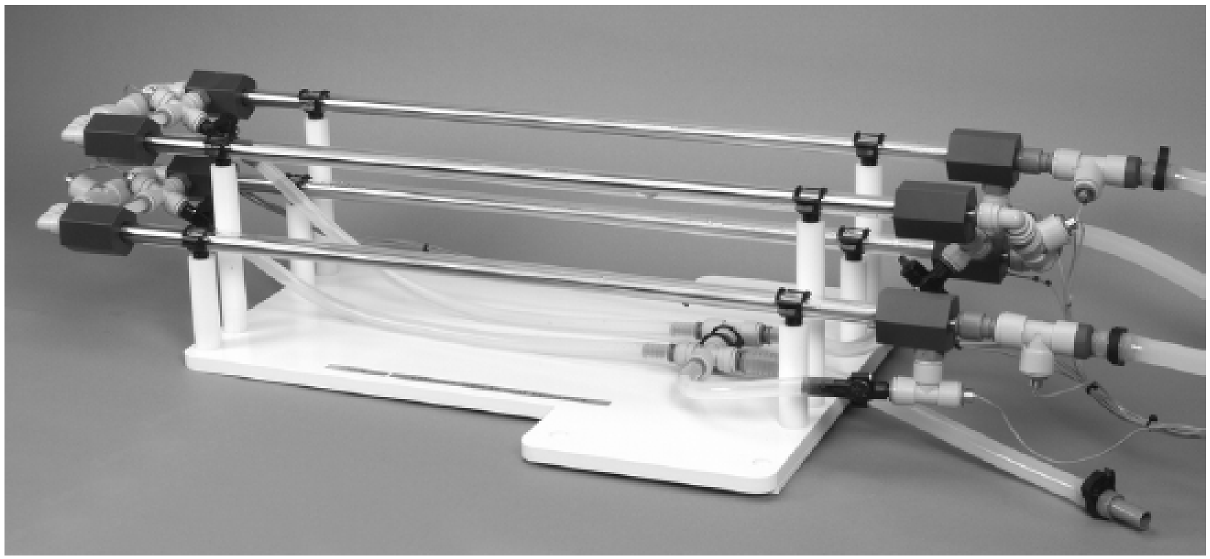
Class Notes L7

Heat Exchanger Theory

Agenda for Today’s Class

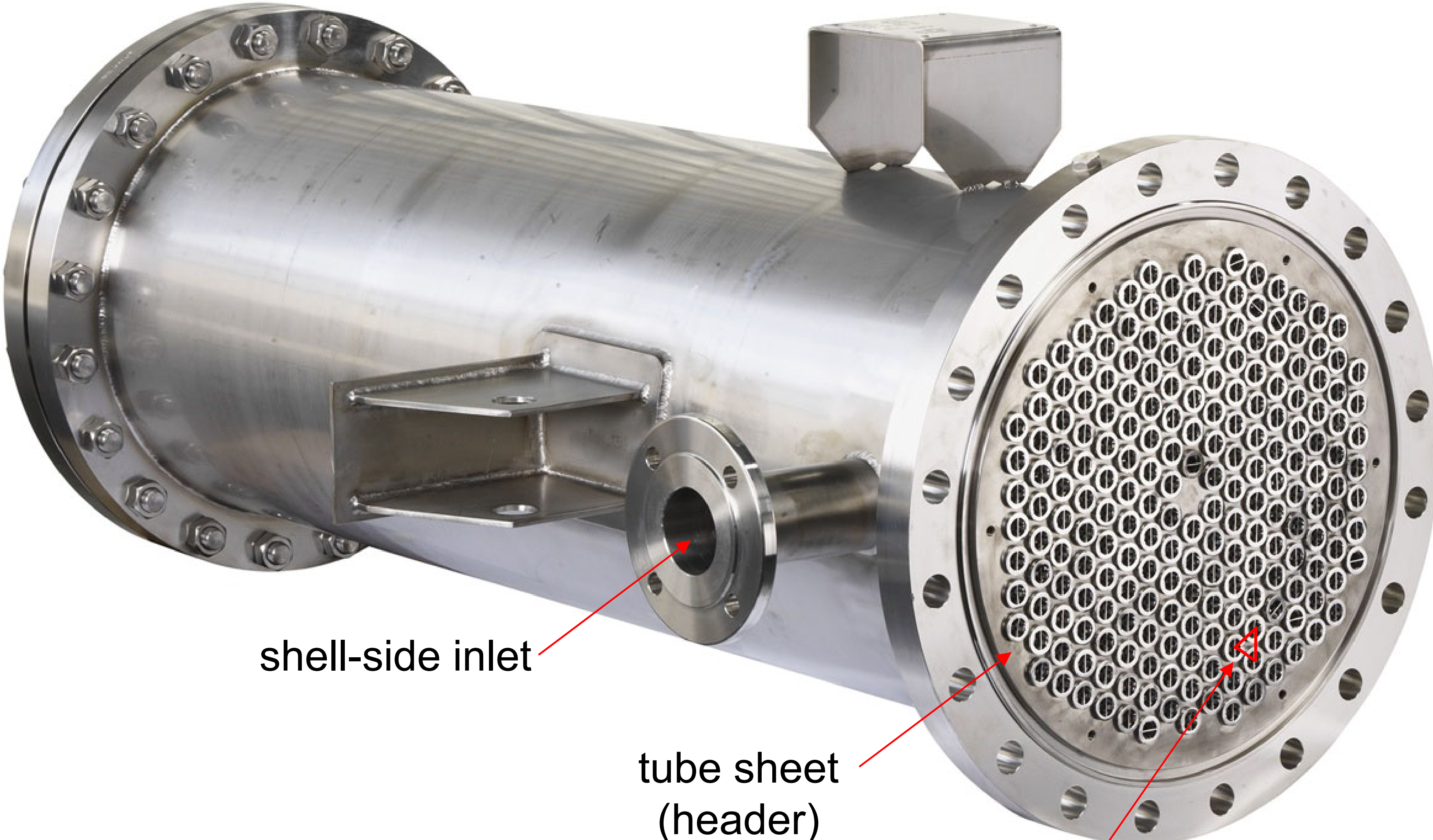
- design and cost factors
- log mean temperature difference
- resistance to heat transfer
- pressure drop (combines L1-6 with CH485)

Simple Tubular Design – CH459, CH485



HT36 EXTENDED TUBULAR HEAT EXCHANGER

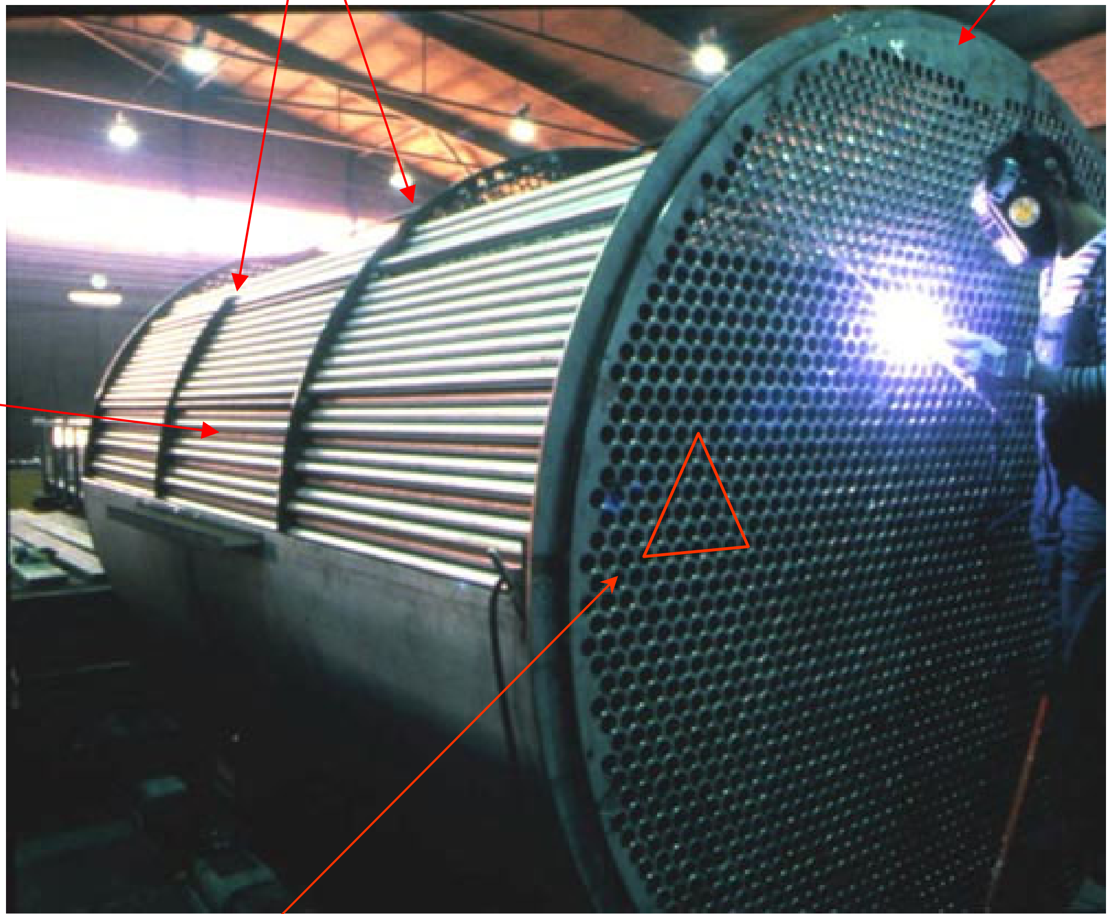
Basic “shell-and-tube” design



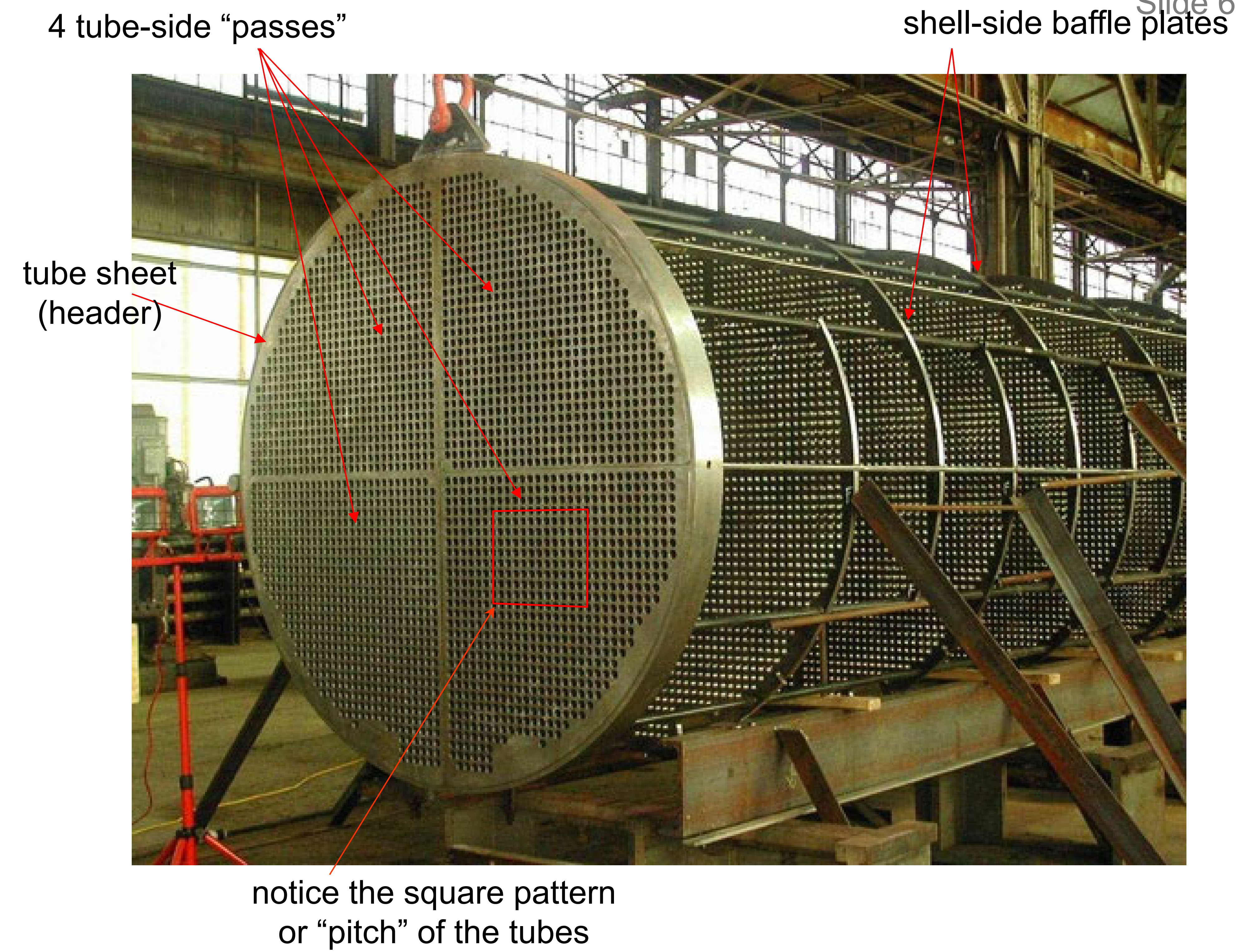
baffle plates

tube sheet

tube bundle



triangular tube “pitch”


4 tube-side "passes"
shell-side baffle plates
tube sheet
(header)
notice the square pattern
or "pitch" of the tubes

Heat exchangers can be quite large…

Cost goes up exponentially with size.

What cost factors are most important?

Heat transfer area and material

p. 682

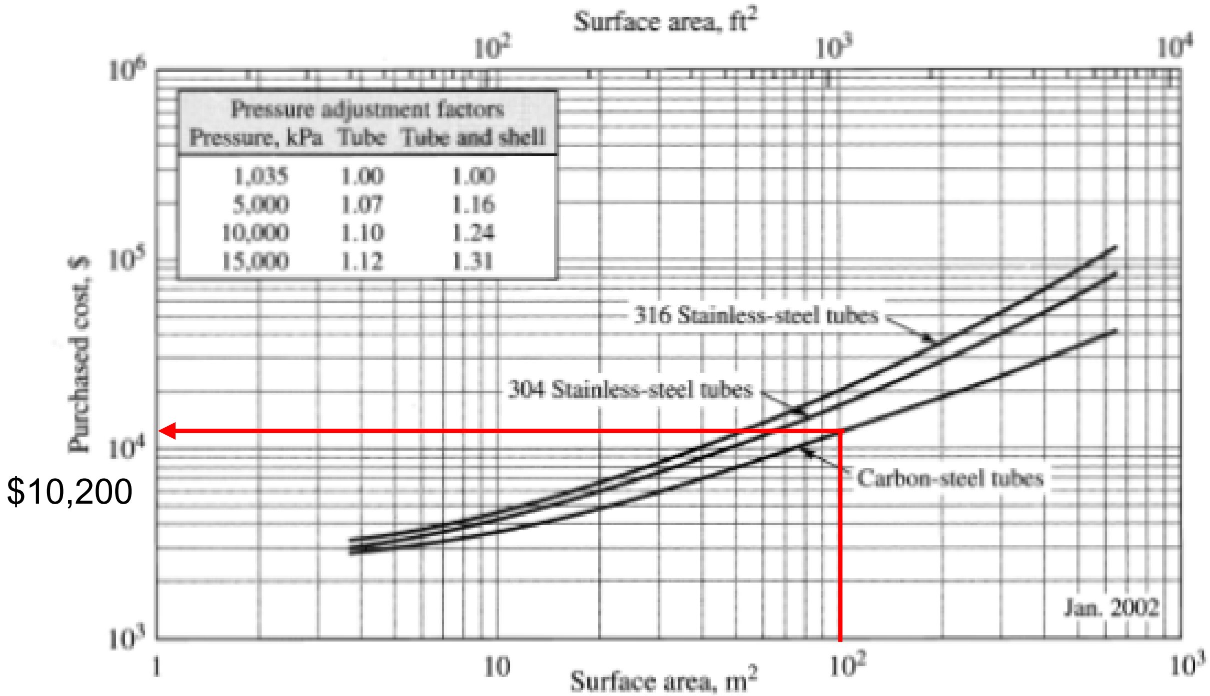


Figure 14-18
Purchased cost of fixed-tube-sheet heat exchangers with 0.019-m ($\frac{3}{4}$-in.) OD × 0.025-m (1-in.) square pitch and 4.88- or 6.10-m (16- or 20-ft) bundles and carbon-steel shell operating at 1035 kPa (150 psia)

carbon steel, 100 m^2: $10,200

¾-in tubes, 1-inch square-pitch fixed tube sheet, 4.88 m or 6.10 m in length at 150 psia

What if we change design to 1.5-inch tubes, 4 m in length, 500 psia, stainless 316L, in Jan 2024?

1.5-in tubes, multiply by 1.37, figure 14-21: $13,974

4-m length, multiply by 1.07, figure 14-22: $14,952

500 psia, multiply by 1.39, figure 14-23: $20,784

Stainless 316L, multiply by 2.2, table 14-8: $45,724

Jan 2024, multiply by 786.6/356.9, CEPCI: $100,774 // ANS

Summary of cost factors and Adjustments

All cost charts located on pages 680 – 692 Very Important!

Basic shell-and-tube heat exchangers, Figures 14-18 and 14-19, page 682

0.019-m x 4.88-m, fixed-tube-sheet, Figure 14-18, p. 682

0.019-m x 4.88-m, floating-head, Figure 14-19, p. 682

Adjustments

Tube diameter, Figure 14-21, p. 683

Tube length, Figure 14-22, p. 684

Pressure, Figure 14-23, p. 684

Material, Table 14-8 and Figure 14-24, p. 685

Today's Class

- design and cost factors
- log mean temperature difference
- resistance to heat transfer
- pressure drop

Basic Heat Exchanger Design - Textbook

Mean overall temperature difference between the fluids.

$$\dot{q} = U \cdot A \cdot \Delta T_{o,m}$$ Eq. 14-5

Linear temperature profiles inside heat exchanger (constant Cp)

$$\dot{q} = U \cdot A \cdot F \cdot \Delta T_{o,\log\ mean}$$

Combine w/ Eq. 14-8

Eq. 14-8

$$\Delta T_{o,m} = F \cdot \Delta T_{o,\ \log\ mean}$$

F accounts for arrangement of exchanger

$\dot{q} =$ rate of heat transfer, kJ/s or kW

$U =$ overall heat transfer coefficient, W/m^2K

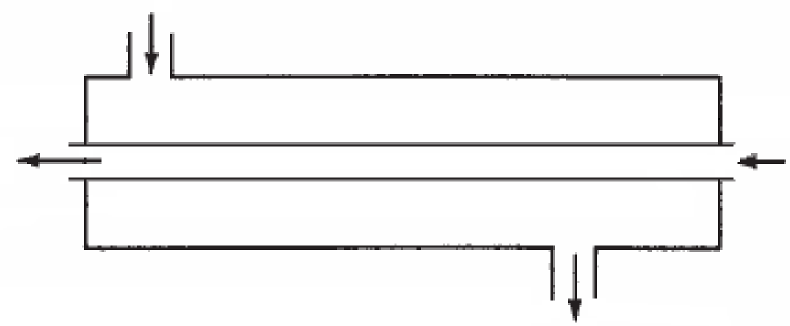
$A =$ area of heat transfer, m^2

$\Delta T =$ mean temperature-difference driving force, K

U is calculated from the resistance model

A is calculated from the geometry

$$A = \pi \cdot D \cdot L \cdot N_t$$



$$\dot{q} = -\left(\dot{m} C_p \Delta T\right)_{hot} = \left(\dot{m} C_p \Delta T\right)_{cold}$$

Conservation of enthalpy
Equation 14-6
Page 645

heat leaving hot side = heat entering cold side

ideal = no heat lost (LHS=RHS)

FE Reference Manual

FE Reference Manual, page 215 (221/500)

Heat Exchangers

The rate of heat transfer in a heat exchanger is

$$\dot{Q} = UAF\Delta T_{lm}, \text{ where}$$

(Manual does not say how to determine F)

A = any convenient reference area (m^2)

F = correction factor for log mean temperature difference for more complex heat exchangers (shell and tube arrangements with several tube or shell passes or cross-flow exchangers with mixed and unmixed flow); otherwise $F = 1$.

U = overall heat-transfer coefficient based on area A and the log mean temperature difference [W/(m^2•K)]

ΔT_{lm} = log mean temperature difference (K)

Log Mean Temperature Difference (LMTD)

For *counterflow* in tubular heat exchangers

$$\Delta T_{lm} = \frac{(T_{Ho} - T_{Ci}) - (T_{Hi} - T_{Co})}{\ln\left(\frac{T_{Ho} - T_{Ci}}{T_{Hi} - T_{Co}}\right)}$$

For *parallel flow* in tubular heat exchangers

$$\Delta T_{lm} = \frac{(T_{Ho} - T_{Co}) - (T_{Hi} - T_{Ci})}{\ln\left(\frac{T_{Ho} - T_{Co}}{T_{Hi} - T_{Ci}}\right)}, \text{ where}$$

ΔT_{lm} = log mean temperature difference (K)

T_{Hi} = inlet temperature of the hot fluid (K)

T_{Ho} = outlet temperature of the hot fluid (K)

T_{Ci} = inlet temperature of the cold fluid (K)

T_{Co} = outlet temperature of the cold fluid (K)

U is calculated from the resistance model

FE Reference Manual, page 216 (222/500)

A is calculated from the geometry

$$A = \pi \cdot D \cdot L \cdot N_t$$

Overall Heat-Transfer Coefficient for Concentric Tube and Shell-and-Tube Heat Exchangers

$$\frac{1}{UA} = \frac{1}{h_i A_i} + \frac{R_{fi}}{A_i} + \frac{\ln\left(\frac{D_o}{D_i}\right)}{2\pi kL} + \frac{R_{fo}}{A_o} + \frac{1}{h_o A_o}, \text{ where}$$

A_i = inside area of tubes (m^2)

A_o = outside area of tubes (m^2)

D_i = inside diameter of tubes (m)

D_o = outside diameter of tubes (m)

h_i = convection heat-transfer coefficient for inside of tubes [W/(m^2•K)]

h_o = convection heat-transfer coefficient for outside of tubes [W/(m^2•K)]

k = thermal conductivity of tube material [W/(m•K)]

R_{fi} = fouling factor for inside of tube [(m^2•K)/W]

R_{fo} = fouling factor for outside of tube [(m^2•K)/W]

Recommend review of heat transfer section, pp. 204-219; very well written.

Log-mean temperature difference

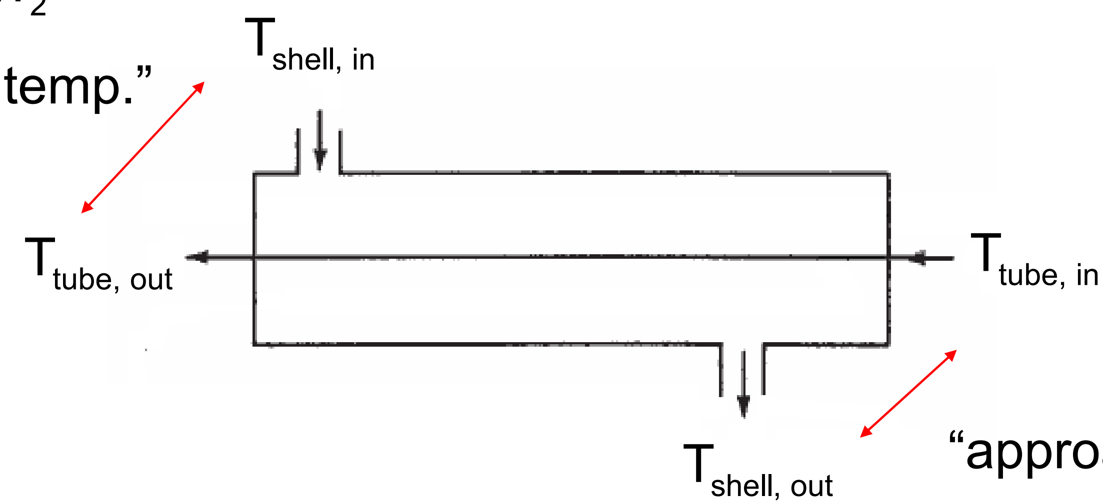
(mean temperature difference)

$$\Delta T_{o,\log\ mean} = \frac{\Delta T_1 - \Delta T_2}{\text{Ln}\left(\Delta T_1 / \Delta T_2\right)}$$

Equation 14-7
Page 645

ΔT_1 or ΔT_2

“approach temp.”



“approach temp.”

ΔT_1 or ΔT_2

$\Delta T_1 > \Delta T_2$

FE Reference Manual, page 215 (221/502)

Parallel (cocurrent) flow

$$\Delta T_{lm} = \frac{\left(T_{Ho} - T_{Co}\right) - \left(T_{Hi} - T_{Ci}\right)}{\text{Ln}\left(\left(T_{Ho} - T_{Co}\right) / \left(T_{Hi} - T_{Ci}\right)\right)}$$

Counterflow (countercurrent) flow

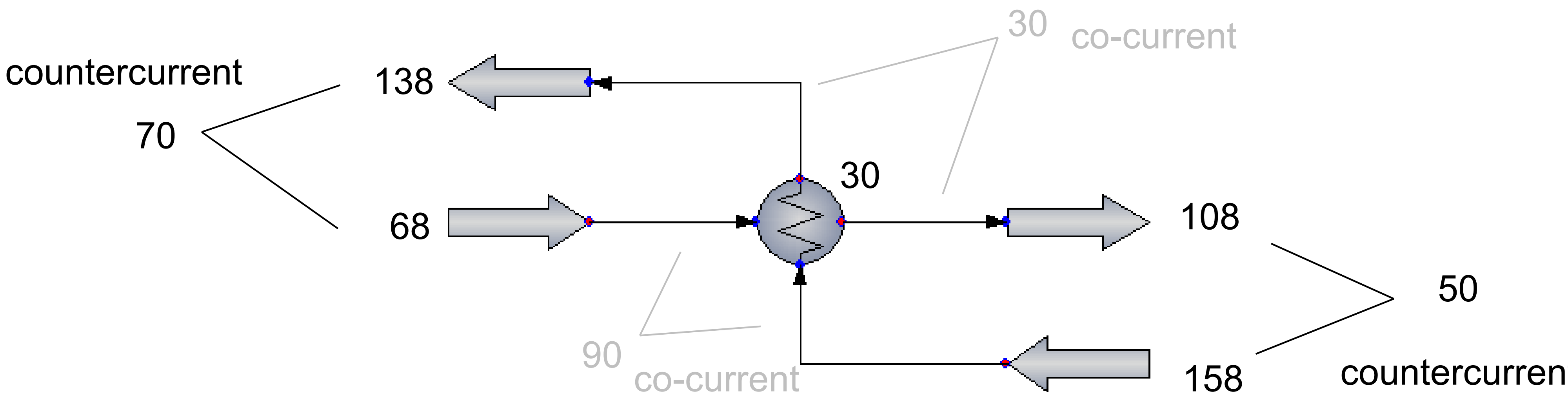
$$\Delta T_{lm} = \frac{\left(T_{Ho} - T_{Ci}\right) - \left(T_{Hi} - T_{Co}\right)}{\text{Ln}\left(\left(T_{Ho} - T_{Ci}\right) / \left(T_{Hi} - T_{Co}\right)\right)}$$

Log mean temperature example 1

Calculate the log mean temperature difference for countercurrent flow.

Temperatures in deg C

$$\Delta T_{lm} = \frac{(T_{Ho} - T_{Ci}) - (T_{Hi} - T_{Co})}{Ln((T_{Ho} - T_{Ci}) / (T_{Hi} - T_{Co}))}$$



$$\Delta T_{o,\log mean} = \frac{\Delta T_1 - \Delta T_2}{\ln[\Delta T_1 / \Delta T_2]}$$

$$\Delta T_{o,\log mean} = \frac{70 - 50}{\ln[70/50]}$$

ΔT_{LM}=59.44

CHEMCAD LMTD is 59.44 ✓

A good check on your answer is the arithmetic mean:

$$\Delta T_{mean} = \frac{\Delta T_1 + \Delta T_2}{2} = \frac{70 + 50}{2} = 60$$

about equal if $\frac{\Delta T_1}{\Delta T_2} \le 2$

Log mean temperature is more complicated in multi-pass heat exchangers

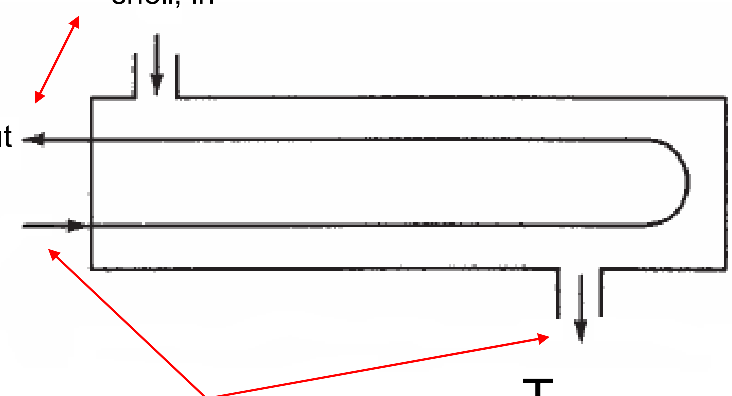
That's why we use "F"

$$\Delta T_{o,\text{log mean}} = \frac{\Delta T_1 - \Delta T_2}{\mathrm{Ln}\left(\Delta T_1 / \Delta T_2\right)}$$

ΔT_1 or ΔT_2

Parallel or Countercurrent?

"approach temp."



"approach temp."

ΔT_1 or ΔT_2

which? $\Delta T_1 > \Delta T_2$

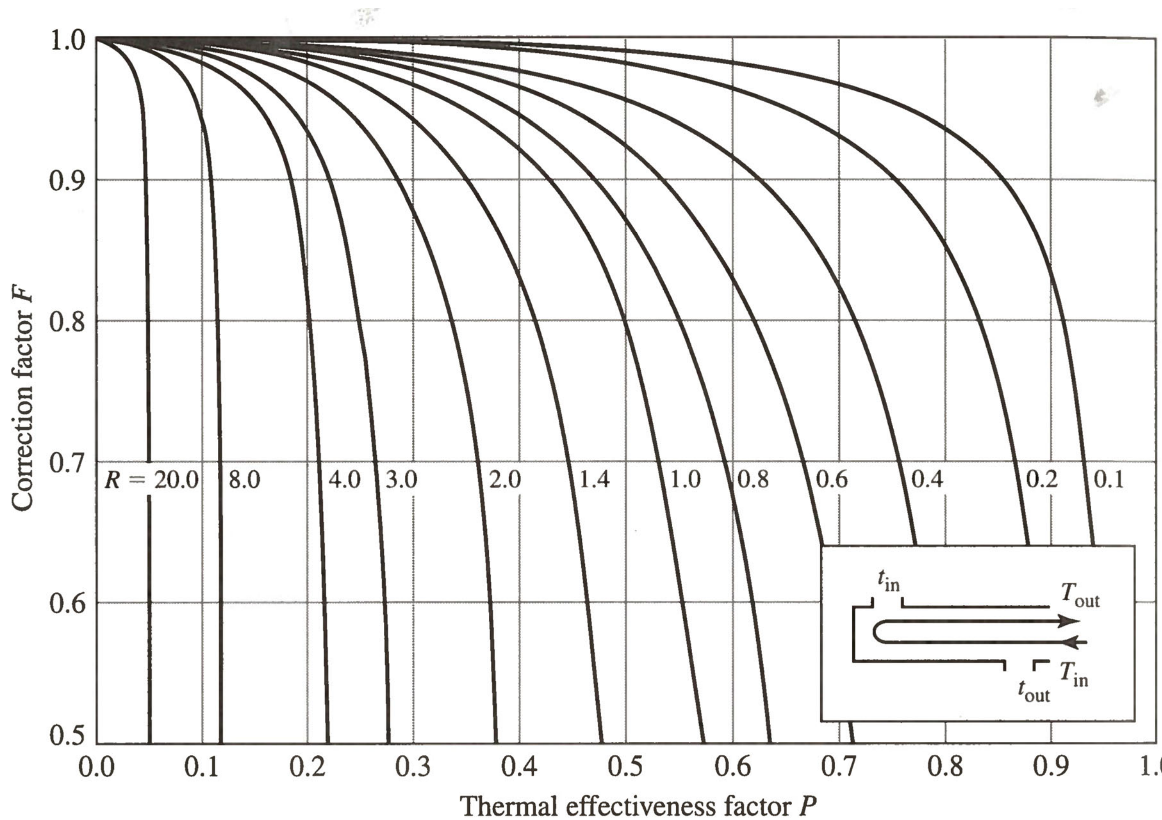
$$\Delta T_{o,m} = F \cdot \Delta T_{o,\text{log mean}}$$

F is a correction factor
for the number of passes

Determining F

2 or more even-numbered tube passes
Figure 14-4, page 648

Figure 14-5, page 649
other contact patterns



$$F = (R^2+1)^{1/2} \cdot \ln\left(\frac{1-P}{1-R\cdot P}\right) \Bigg/ (R-1)\cdot \ln\left(\frac{2-P\cdot\left[(R+1)-(R^2+1)^{1/2}\right]}{2-P\cdot\left[(R+1)+(R^2+1)^{1/2}\right]}\right)$$

$$P = \frac{T_{c,out} - T_{c,in}}{T_{h,in} - T_{c,in}} \qquad R = \frac{T_{h,in} - T_{h,out}}{T_{c,out} - T_{c,in}} = \frac{\left(\dot{m}\cdot C_p\right)_c}{\left(\dot{m}\cdot C_p\right)_h}$$

<u>Good Rule of Thumb:</u>

F > 0.85 desirable

0.7 < F < 0.85 marginal

F < 0.7 impractical

Log mean temperature

Example 2

$$T_{h,\ in} = 158\ ^oC$$

$$T_{h,\ out} = 68\ ^oC$$

$$T_{c,\ in} = 68\ ^oC$$

$$T_{c,\ out} = 108\ ^oC$$

2 tube passes,
1 shell pass

$$\Delta T_1 = 158 - 68 = 90\ ^oC$$

$$\Delta T_2 = 138 - 108 = 30\ ^oC$$

$$\Delta T_{o,\ log\ mean} = \frac{\Delta T_1 - \Delta T_2}{\ln\left(\Delta T_1 / \Delta T_2\right)} = \frac{90 - 30}{\ln\left(90/30\right)} = 54.61$$

Log mean temperature

Example 2 (continued)

$T_{h,\ in} = 158\ ^oC$

$T_{h,\ out} = 68\ ^oC$

$T_{c,\ in} = 68\ ^oC$

$T_{c,\ out} = 108\ ^oC$

2 tube passes,
1 shell pass

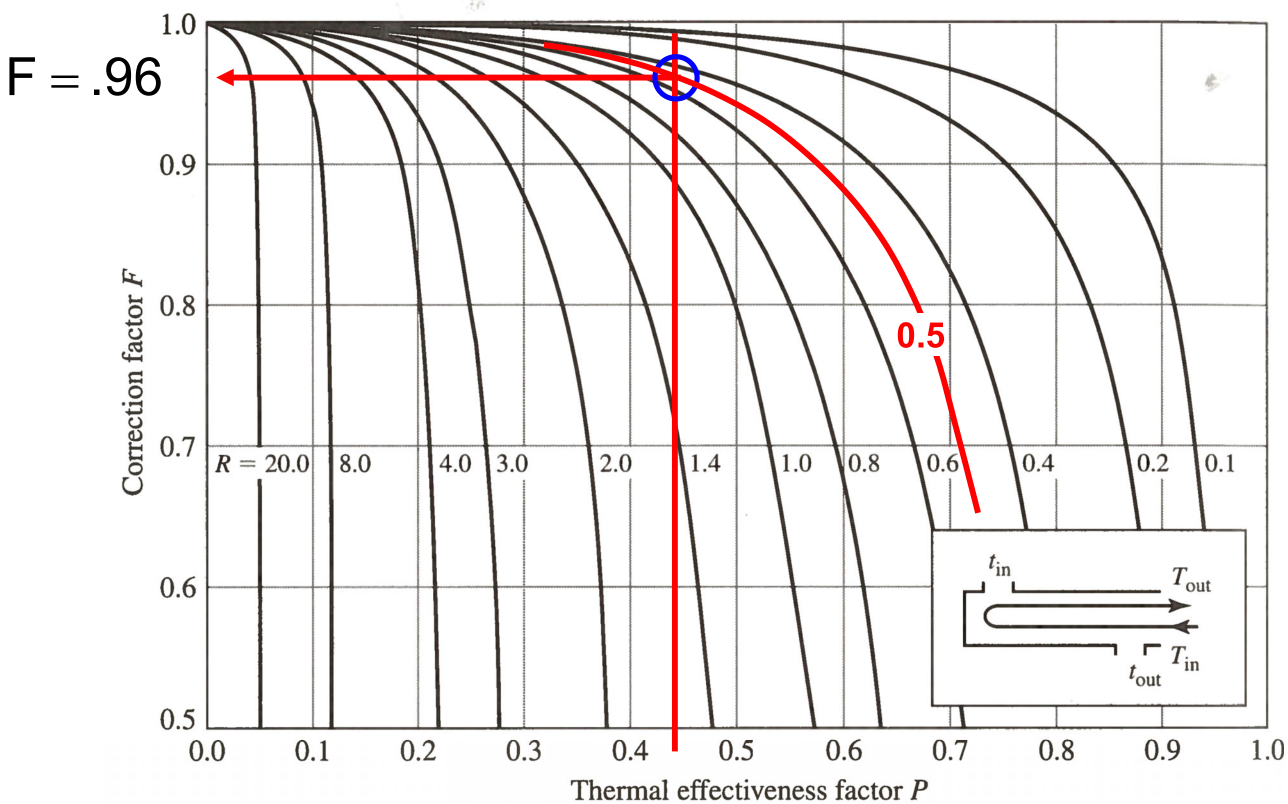
$$R = \frac{T_{h,in} - T_{h,out}}{T_{c,out} - T_{c,in}} = \frac{158 - 138}{108 - 68} = 0.5$$

$$P = \frac{108 - 68}{158 - 68} = 0.44$$

$$F = \frac{(R^2 + 1)^{1/2} \cdot \ln\left(\frac{1 - P}{1 - R \cdot P}\right)}{(R - 1) \cdot \ln\left(\frac{2 - P \cdot \left[(R + 1) - (R^2 + 1)^{1/2}\right]}{2 - P \cdot \left[(R + 1) + (R^2 + 1)^{1/2}\right]}\right)} = 0.961$$

$$\Delta T_{o,m} = F \cdot \Delta T_{o,log\ mean} = .961 \cdot 54.61 = 52.41\ ^oC$$

How to read the chart…



$P = .44 \qquad R = .5$

$$\Delta T_{o,m} = F \cdot \Delta T_{o,\log\ mean} = .96 \cdot 54.61 = 5\underline{2}.42\ ^{o}C$$

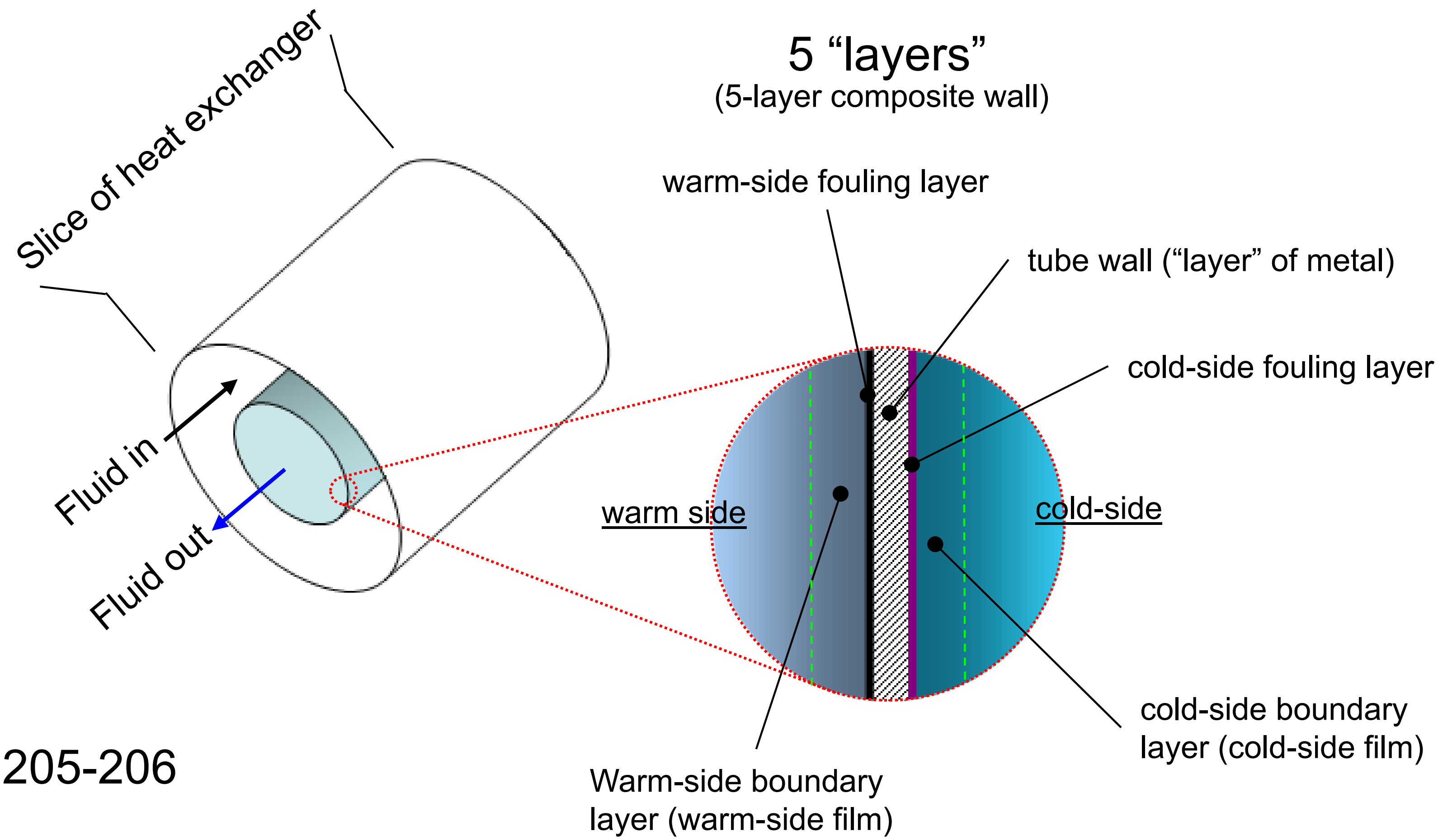
CHEMCAD calculates the correction factors for us

Does not appear in FE manual (F=1)

Today's Class

- design and cost factors
- log mean temperature difference
- resistance to heat transfer
- pressure drop

Thermal Resistances in Series Model



FE Manual, pp. 205-206

Total resistance = sum of resistances of individual layers

$$R_{tot} = \frac{1}{U} = \underbrace{\left(\frac{1}{h_{warm}}\right)}_{\text{warm-side film}} + \underbrace{\left(R_{warm,fouling}\right)}_{\text{warm-side fouling}} + \underbrace{\left(\frac{x_{wall}}{k_w}\right)}_{\text{wall}} + \underbrace{\left(R_{cold,fouling}\right)}_{\text{cold-side fouling}} + \underbrace{\left(\frac{1}{h_{cold}}\right)}_{\text{cold-side film}}$$

Different forms for total resistance

Tube, average: $$R_{tot} = \frac{1}{U} = \frac{1}{h_h} + R_{h,f} + \frac{x}{k_w} + R_{c,f} + \frac{1}{h_c}$$ Eq. 14-2, p. 644

Tube, outside: $$\frac{1}{U_o} = \frac{A_o}{h_i \cdot A_i} + \frac{A_o}{h_{i,d} \cdot A_i} + \frac{A_o \cdot x_w}{k_w \cdot A_{m,w}} + \frac{1}{h_o} + \frac{1}{h_{o,d}}$$ Eq. 14-4, p. 645

Tube, inside: $$\frac{1}{U_i} = \frac{1}{h_i} + \frac{1}{h_{i,d}} + \frac{A_i \cdot x_w}{k_w \cdot A_{m,w}} + \frac{A_i}{h_o \cdot A_o} + \frac{A_i}{h_{o,d} \cdot A_o}$$ Eq. 14-4a, p. 645

FE Reference Manual,
page 205 (211/502)

"<u>Overall Heat Transfer Coefficient for Concentric Tube and Shell-and-Tube Heat Exchangers</u>"

$A_i = \pi \cdot D_i \cdot L$ (inside area)
$A_o = \pi \cdot D_o \cdot L$ (outside area)
L = length

$$\frac{1}{UA} = \frac{1}{h_i \cdot A_i} + \frac{R_{fi}}{A_i} + \frac{\ln\left(\frac{D_o}{D_i}\right)}{2\pi k L} + \frac{R_{fo}}{A_o} + \frac{1}{h_o A_o}$$

$A = A_i$ or A_o

Needed for problem 14-2

Local Heat Transfer Coefficients

FE Prep: Heat Transfer Section, FE Manual, pp. 204-219, comprehensive

Laminar (viscous) fluid flow inside tubes, Re<2100 :

$$\frac{h_i \cdot D_i}{k} = 1.86 \cdot \left(\frac{D_i \cdot G_i}{\mu}\right)^{1/3} \cdot \left(\frac{C_p \cdot \mu}{k}\right)^{1/3} \cdot \left(\frac{D_i}{L}\right)^{1/3} \cdot \left(\frac{\mu}{\mu_w}\right)^{.14}$$

14-17, p. 657

FE Reference Manual, page 210 (216/502)

$$h_i = 1.86 \cdot \left(\frac{4 \cdot \dot{m}_i \cdot C_p}{\pi \cdot k \cdot L}\right)^{1/3} \cdot \left(\frac{\mu}{\mu_w}\right)^{.14}$$

$$G = \text{mass velocity in } \frac{\text{kg}}{\text{m}^2\,\text{sec}} = \underbrace{\text{velocity}}_{\frac{\text{m}}{\text{sec}}} \times \underbrace{\text{density}}_{\frac{\text{kg}}{\text{m}^3}} = v \times \rho$$

Turbulent, fully developed fluid flow inside tubes, Re>10000, 0.7<Pr<160, and L/D_i>10

14-18, p. 657

book typo here

$$\underbrace{\frac{h_i \cdot D_i}{k}}_{\text{Nusselt number}} = .023 \cdot \underbrace{\left(\frac{D_i \cdot G_i}{\mu}\right)}_{\text{Reynolds number}}{}^{4/5} \cdot \underbrace{\left(\frac{C_p \cdot \mu}{k}\right)}_{\text{Prandtl number}}{}^{1/3} \cdot \left(\frac{\mu}{\mu_w}\right)^{.14}$$

Dittus-Boelter Equation
FE Reference Manual, page 211 (217/502)

Local Heat Transfer Coefficients

Guielinski-Hewitt Correlation

2300<Re<100,000 and 0.6<Pr<6000

book

Equations 14-19 and 14-19a, page 658

$$\frac{h_i \cdot D_i}{k} = \frac{(f_D / 8) \cdot (Re - 1000) \cdot Pr}{\left[1 + 12.7 \cdot (f_D / 8)^{1/2}\right] \cdot (Pr^{2/3} - 1)} \cdot \left[1 - \left(\frac{D_i}{L}\right)^{2/3}\right]$$

Apparent typo since Pr must be > 1

(If Pr<1 then Nu<0)

$$f_D = \frac{1}{(1.82 \cdot \log(Re) - 1.64)^2}$$

More accurate and more widely applicable

Local Heat Transfer Coefficients

Outside tubes (Churchill-Bernstein Correlation)

$$\frac{h_o \cdot D_o}{k} = 0.3 + \frac{0.62 \cdot Re^{1/2} \cdot Pr^{1/3}}{[1 + (0.4 / Pr)^{2/3}]^{1/4}} \cdot \left[1 + \left(\frac{Re}{28,200} \right)^{5/8} \right]^{4/5}$$

14-21, p. 659

Multiple tubes in parallel:

$$\frac{h_o \cdot D_o}{k} = a \cdot Re^m \cdot Pr^{0.34} \cdot F_1 \cdot F_2$$

14-22, p. 660

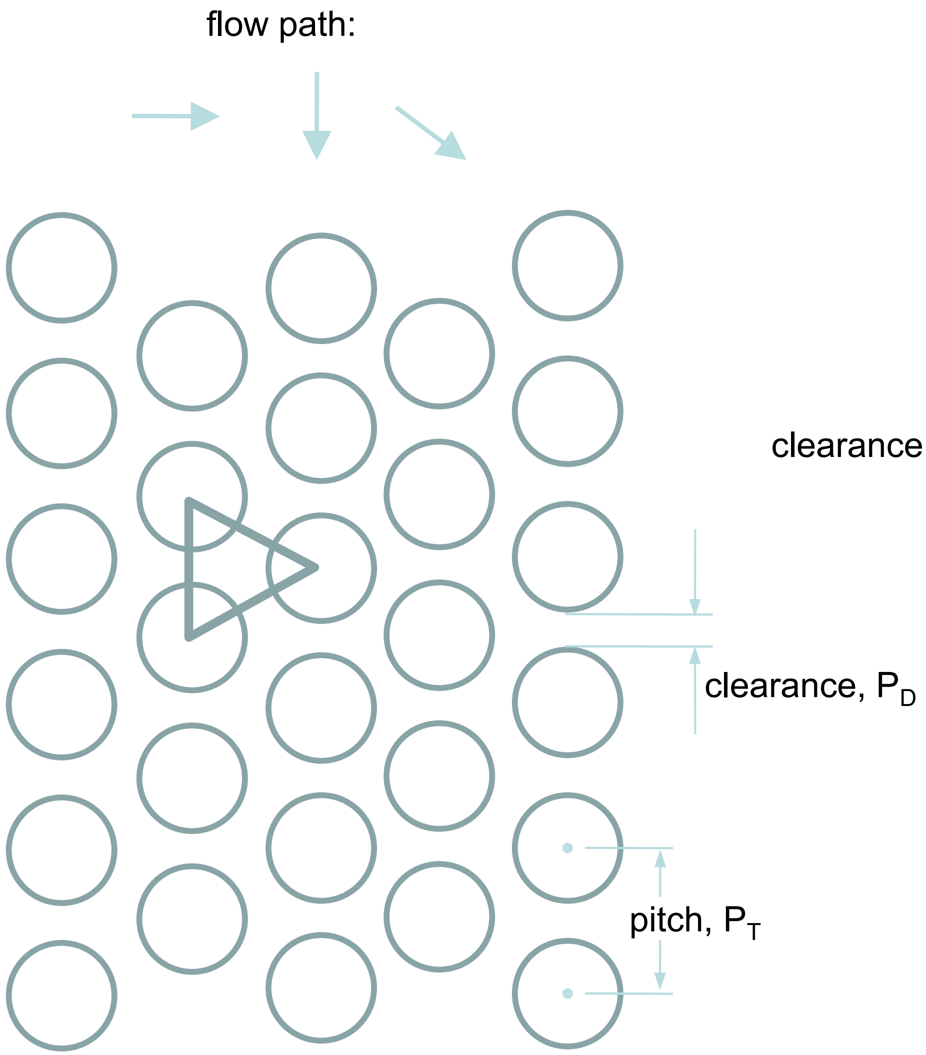
$$F_1 = \left(\frac{Pr_b}{Pr_w} \right)^{0.26}$$

a and m are found in Table 14-1

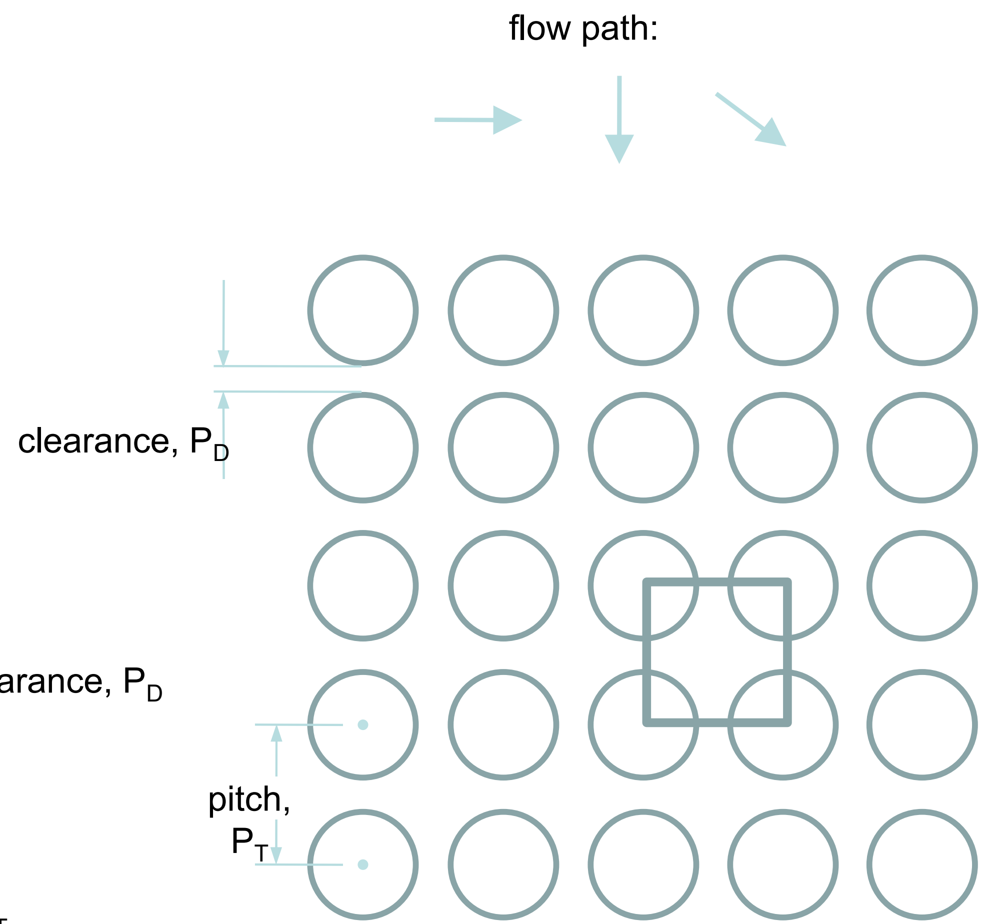
F_2 is found in Table 14-2

(Not found in FE Manual – inside tubes only)

Tube Bank Patterns in Shell-and-Tube Heat Exchangers



triangular pitch



square pitch

Film Coefficients for Banks of Tubes

Shell-side Nusselt number:

$$Nu = \frac{h_o \cdot D_o}{k} = a \cdot Re^m \cdot Pr^{0.34} \cdot F_1 \cdot F_2$$

Eq. 14-22
p. 660

$$F_1 = \left(\frac{Pr_{bulk}}{Pr_{wall}}\right)^{.26} \quad \text{for } Pr < 600$$

Eq. 14-22b
p. 660

Shell-side Reynolds number:

$$Re = \frac{D_o \cdot v_{max} \cdot \rho}{\mu}$$

Eq. 14-22a
p. 660

$$v_{max} = \frac{\dot{m}_T}{\rho \cdot N_T \cdot (P_T - D_O) \cdot L}$$

Eq. 14-22c
p. 661

Table 14-2. Correction factor F_2 for eq 14-22

Number of tube rows	F_2, inline banks (Re>2100)	F_2, staggered banks (Re>100)
3	0.86	0.85
4	0.90	0.90
6	0.94	0.95
8	0.98	0.985
10+	0.99	0.99

Table 14-1. Values of a and m for eq. 14-22

Re	Inline banks $1.2 < P_T/D_o < 4$		Staggered banks (Re>100)	
	a	m	a	m
10-300	0.742	0.431	1.309	0.360
$300\text{-}2\times10^5$	0.211	0.651	0.273	0.635
$2\times10^5\text{-}2\times10^6$	0.116	0.700	0.124	0.700

Condensers

Horizontal tube, outside laminar condensation, coolant inside tubes:

$$h_o = 0.725 \cdot \left[\frac{k_L^3 \cdot \rho_L \cdot \left(\rho_L - \rho_V\right) \cdot g \cdot \lambda_C}{\mu_L \cdot D_o \cdot \left(T_{sat} - T_{wall}\right)} \right]$$

k_L, ρ_L, and μ_L refer to liquid condensate
ρ_V refers to vapor
g is gravitational acceleration
λ_C is the heat of condensation
T_{sat} is the temperature of the condensate
T_{wall} is the wall temperature

$$h_o = \frac{3100}{D_o^{1/4} \cdot \left(T_{sat} - T_{wall}\right)^{1/3}} \quad \text{(steam)}$$

Tube bundles:

$$h_N = h_o \cdot N_V^{-1/4}$$

accounts for liquid dripping from tubes to lower tubes from Perry's Handbook

TABLE 11-3 Typical Overall Heat-Transfer Coefficients in Tubular Heat Exchangers

U = Btu/(°F · ft² · h)

Shell side	Tube side	Design U	Includes total dirt
Liquid-liquid media			
Aroclor 1248	Jet fuels	100–150	0.0015
Cutback asphalt	Water	10–20	.01
Demineralized water	Water	300–500	.001
Ethanol amine (MEA or DEA) 10–25% solutions	Water or DEA, or MEA solutions	140–200	.003
Fuel oil	Water	15–25	.007
Fuel oil	Oil	10–15	.008
Gasoline	Water	60–100	.003
Heavy oils	Heavy oils	10–40	.004
Heavy oils	Water	15–50	.005
Hydrogen-rich reformer stream	Hydrogen-rich reformer stream	90–120	.002
Kerosene or gas oil	Water	25–50	.005
Kerosene or gas oil	Oil	20–35	.005
Kerosene or jet fuels	Trichlorethylene	40–50	.0015
Jacket water	Water	230–300	.002
Lube oil (low viscosity)	Water	25–50	.002
Lube oil (high viscosity)	Water	40–80	.003
Lube oil	Oil	11–20	.006
Naphtha	Water	50–70	.005
Naphtha	Oil	25–35	.005
Organic solvents	Water	50–150	.003
Organic solvents	Brine	35–90	.003
Organic solvents	Organic solvents	20–60	.002
Tall oil derivatives, vegetable oil, etc.	Water	20–50	.004
Water	Caustic soda solutions (10–30%)	100–250	.003
Water	Water	200–250	.003
Wax distillate	Water	15–25	.005
Wax distillate	Oil	13–23	.005
Condensing vapor-liquid media			
Alcohol vapor	Water	100–200	.002
Asphalt (450°F.)	Dowtherm vapor	40–60	.006
Dowtherm vapor	Tall oil and derivatives	60–80	.004
Dowtherm vapor	Dowtherm liquid	80–120	.0015
Gas-plant tar	Steam	40–50	.0055
High-boiling hydrocarbons V	Water	20–50	.003
Low-boiling hydrocarbons A	Water	80–200	.003
Hydrocarbon vapors (partial condenser)	Oil	25–40	.004
Organic solvents A	Water	100–200	.003
Organic solvents high NC, A	Water or brine	20–60	.003
Organic solvents low NC, V	Water or brine	50–120	.003
Kerosene	Water	30–65	.004
Kerosene	Oil	20–30	.005
Naphtha	Water	50–75	.005
Naphtha	Oil	20–30	.005
Stabilizer reflux vapors	Water	80–120	.003
Steam	Feed water	400–1000	.0005
Steam	No. 6 fuel oil	15–25	.0055
Steam	No. 2 fuel oil	60–90	.0025
Sulfur dioxide	Water	150–200	.003
Tall-oil derivatives, vegetable oils (vapor)	Water	20–50	.004
Water	Aromatic vapor-stream azeotrope	40–80	.005
Gas-liquid media			
Air, N_2, etc. (compressed)	Water or brine	40–80	.005
Air, N_2, etc., A	Water or brine	10–50	.005
Water or brine	Air, N_2 (compressed)	20–40	.005
Water or brine	Air, N_2, etc., A	5–20	.005
Water	Hydrogen containing natural-gas mixtures	80–125	.003
Vaporizers			
Anhydrous ammonia	Steam condensing	150–300	.0015
Chlorine	Steam condensing	150–300	.0015
Chlorine	Light heat-transfer oil	40–60	.0015
Propane, butane, etc.	Steam condensing	200–300	.0015
Water	Steam condensing	250–400	.0015

NC = noncondensable gas present.
V = vacuum.
A = atmospheric pressure.
Dirt (or fouling factor) units are (h · ft² · °F)/Btu.
To convert British thermal units per hour-square foot-degrees Fahrenheit to joules per square meter-second-kelvins, multiply by 5.6783; to convert hours per square foot-degree Fahrenheit-British thermal units to square meters per second-kelvin-joules, multiply by 0.1761.

See Tables 14-3 to 14.5, pages 661-663 in PTW

Today's Class

- design and cost factors
- log mean temperature difference
- resistance to heat transfer
- pressure drop

Needed for problem 14-2

Pressure Drop – Tube-side

pages 664-665 and footnote on page 665

$$\Delta p_i = \frac{2 \cdot \beta_i \cdot f_i \cdot G_i^2 \cdot L \cdot n_p}{\rho_i \cdot D_i \cdot \Phi_i}$$

Eq 14-23, p. 664

$$f_i = \begin{cases} 0.079\,Re^{-0.25} & \text{for } Re \le 2100 \\ 0.046\,Re^{-0.20} & \text{for } Re > 2100 \end{cases}$$

Eq 14-23a,b

$$\Phi_i = \begin{cases} 1.10\left(\mu_i / \mu_w\right)^{0.25} & \text{for } Re \le 2100 \\ 1.02\left(\mu_i / \mu_w\right)^{0.14} & \text{for } Re > 2100 \end{cases}$$

$$\beta_i = \begin{cases} 1 + \dfrac{F_e + F_c + F_R}{2 \cdot f_i \cdot G_i^2 \cdot L / (\rho_i^2 \cdot D_i \cdot \Phi_i)} \\ 1 + \dfrac{0.51 \cdot K_1 \cdot n_p \cdot \Delta T_{f,i} \cdot (\mu_i / \mu_w)^{0.28}}{(T_{i,\,in} - T_{i,\,out}) \cdot Pr^{2/3}} \end{cases}$$

Eq 14-23c,d, p. 665

$$F_e = \frac{(V_1 - V_2)^2}{2}$$ expansion

see T12.1, pp. 490-491 for V_1, V_2, and K_c

$$F_c = \frac{K_c V_2^{\,2}}{2}$$ contraction

$$F_r = \frac{0.5 V_2^{\,2} \cdot (n_p - 1)}{2 \cdot n_p}$$ flow reversal

$$K_1 = (1 - S_i / S_H)^2 + K_c + 0.5 \cdot (n_p - 1) / n_p$$

$$S_i / S_H = \text{area ratio (cr-sect, tubes to header)}$$

$f_i =$ fanning friction factor based on average T

$\Phi_i =$ correction factor for nonisothermal flow

$\mu_i =$ viscosity at average T

$\mu_w =$ viscosity at wall T

$n_p =$ number of tubes

$L =$ length

$\beta_i =$ correction factor for sudden expansion and contraction

G_i = mass velocity, kg/s·m^2

Pressure Drop – Shell-side

$$\Delta p_o = \frac{2 \cdot B_o \cdot f' \cdot N_{tr} \cdot G_s^2}{\rho_o}$$

$B_o =$ num. of tube crossings

$N_{tr} =$ num. of tube rows

$$f' = b_o \left(\frac{D_o \cdot G_s}{\mu_o} \right)^{-0.15} \quad \text{for } 2000 < Re < 40000$$

$$b_o = \begin{cases} 0.44 + \dfrac{0.08 \cdot x_L}{(x_T - 1)^{(0.43-1.13)/x_L}} & \text{for inline tubes} \\ 0.23 + \dfrac{0.11}{(x_T - 1)^{1.08}} & \text{for staggered tubes} \end{cases}$$

x_T is "pitch" or center to center distance

x_L is ratio of pitch to outside tube diameter

Homework

Problem Set 4

Problems 14-2 and 14-9, Due Thursday 2 Feb

Problem 14-2

At an average film temperature of 350 K, what are the individual heat transfer coefficients when the fluid flowing in a 0.0254-m inside diameter tube is air, water or oil? Each fluid in this comparison exhibits a Reynolds number of 5×10^4. How would the pressure drop vary for each fluid? The relevant properties of the three fluids at 350 K are listed in the table below.

	Air	Water	Oil
Density, kg/m^3	.955	973	854
Viscosity, Pa0087·s	2×10^{-5}	3.72×10^{-4}	3.56×10^{-2}
Thermal Conductivity, W/m·K	0.030	0.668	0.138
Heat Capacity, J/kg·K	1050	4190	2116