CH402 Chemical Engineering Process Design

Class Notes L7

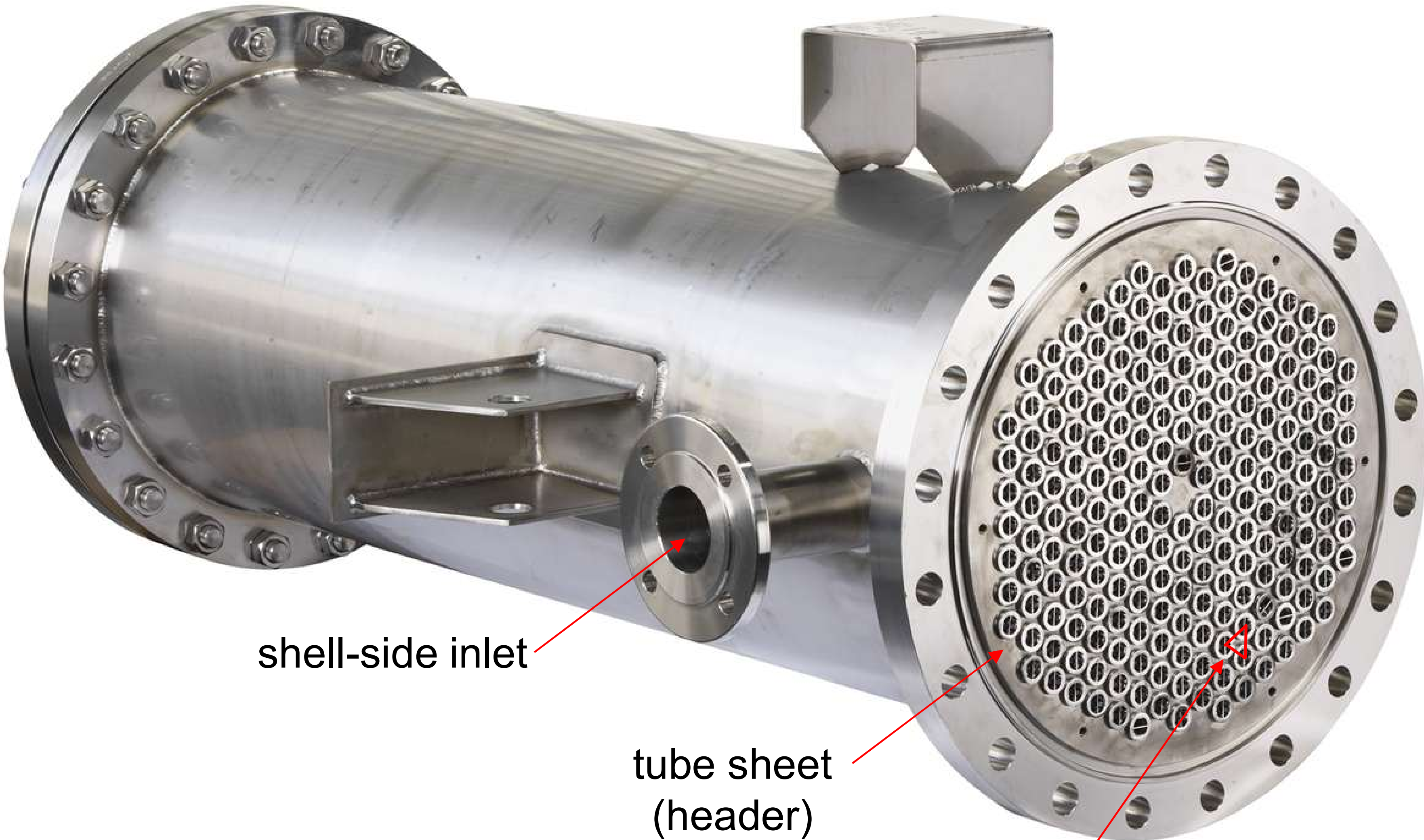
Heat Exchanger Theory

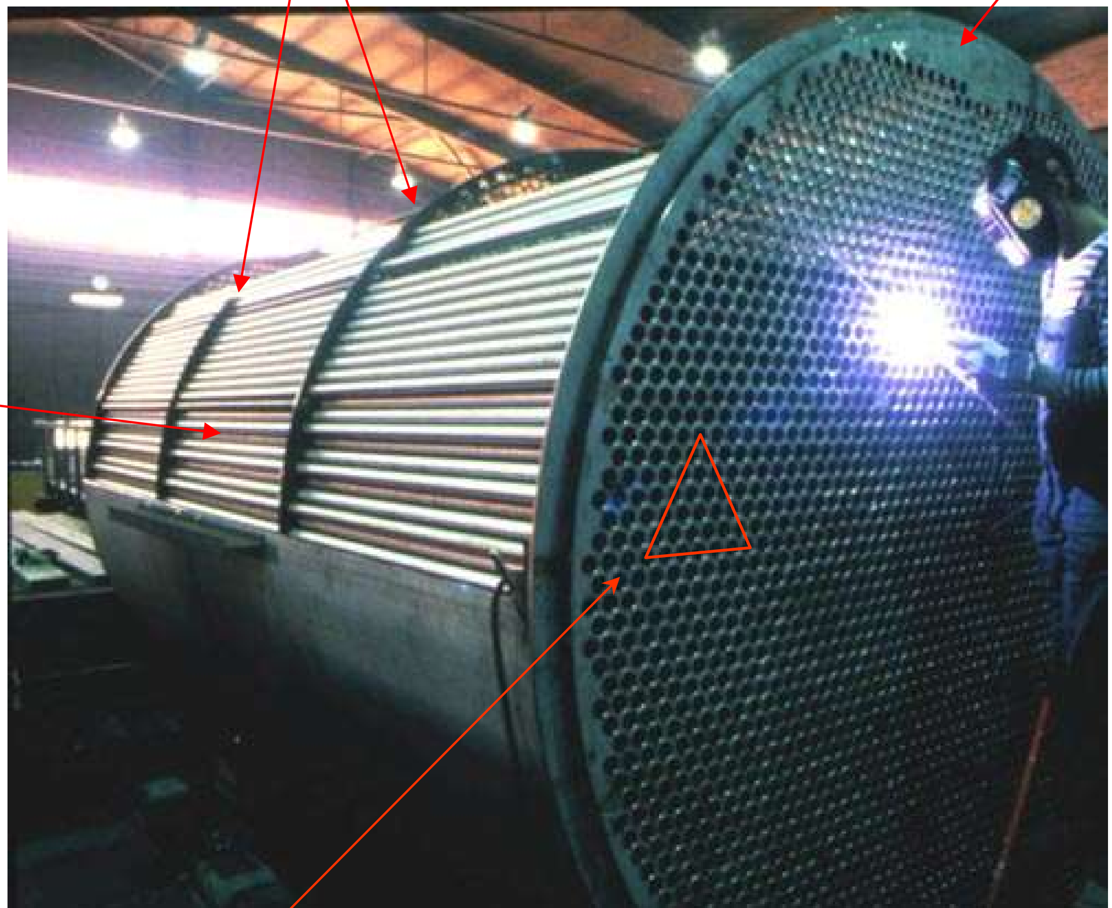
FEE Reference Manual – pp. 214-216

Agenda for Today's Class

- design and cost factors (slides 3-7)
- heat transfer and LMTD (slides 8-11)
- resistance to heat transfer (slides 12-18)
- pressure drop (slides 19-21)

Basic “shell-and-tube” design




baffle plates
tube sheet
tube bundle
triangular tube “pitch”

4 tube-side “passes”

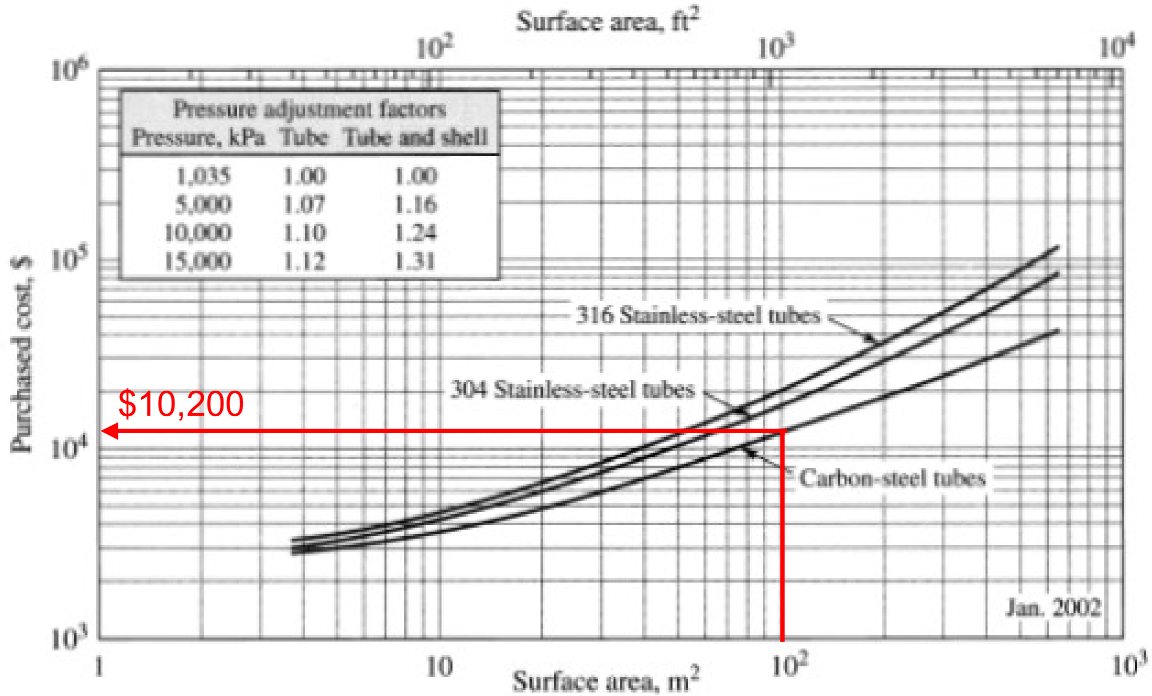
shell-side baffle plates

tube sheet
(header)

notice the square pattern
or “pitch” of the tubes

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 in Jan 2002

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

What if we change design to 1.5-inch tubes, 4 m in length, 500 psia, stainless 316L, in Jan 2026?

Apply correction factors:
Diameter, figure 14-21
Length, figure 14-22
Pressure, figure 14-23
Stainless, figure 14-24
Other materials, table 14-8
(pages 683-685)

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 2026, multiply by 815.7 / 356.9, CEPCI: $104,502 // 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

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FEE Reference Manual – pp. 214-216

Recommend review of entire heat transfer section pp. 203-218; very well written.

Heat Exchangers

The rate of heat transfer associated with either stream in a heat exchanger in which incompressible fluid or ideal gas with constant specific heats flows is

$$\dot{Q} = \dot{m} c_p \left(T_{\text{exit}} - T_{\text{inlet}}\right)$$

c_p = specific heat (at constant pressure)

$\dot{m}$ = mass flow rate

The rate of heat transfer in a heat exchanger is

$$\dot{Q} = UAF \Delta T_{lm}$$

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 [$W/(m^2 \cdot K)$]

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

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

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

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 [$W/(m^2 \cdot K)$]

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

k = thermal conductivity of tube material [$W/(m \cdot K)$]

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

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

A is calculated from the geometry

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

Log Mean Temperature Difference (LMTD)

For *counterflow* in tubular heat exchangers

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

For *parallel flow* in tubular heat exchangers

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

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)

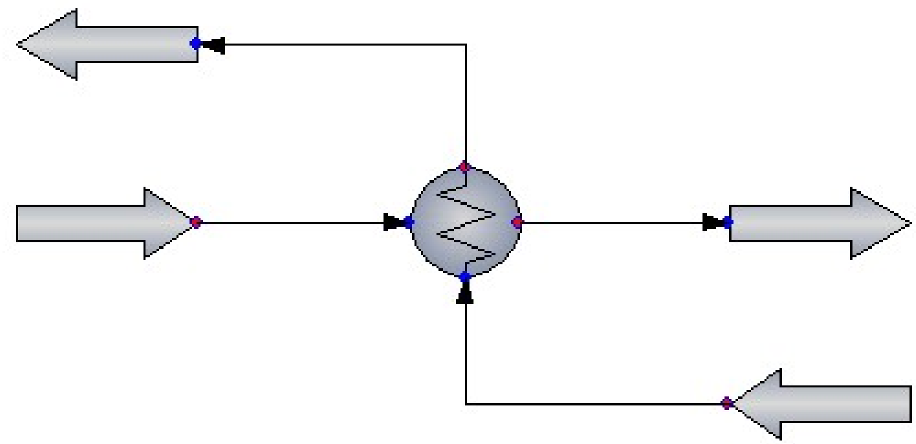
T_{Ci} = inlet temperature of the cold fluid (K)

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

Log mean temperature difference - example

Calculate the log mean temperature difference for countercurrent and co-current flow.

FE Reference Manual, page 214



countercurrent (counterflow)

$$\Delta T_{lm} = \frac{(T_{Ho} - T_{Ci}) - (T_{Hi} - T_{Co})}{Ln\left((T_{Ho} - T_{Ci}) / (T_{Hi} - T_{Co})\right)} = \frac{(138-68)-(158-108)}{Ln\left((138-68)/(158-108)\right)} = 59.44°C$$

for comparison, average temperature difference driving force $\Longrightarrow$ $\Delta T_{mean} = \frac{70+50}{2} = 60$

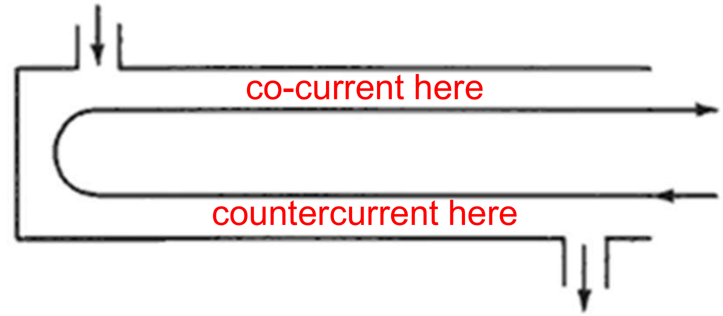
co-current (parallel flow)

$$\Delta T_{lm} = \frac{(T_{Ho} - T_{Co}) - (T_{Hi} - T_{Ci})}{Ln\left((T_{Ho} - T_{Co}) / (T_{Hi} - T_{Ci})\right)} = \frac{(138-108)-(158-68)}{Ln\left((138-108)/(158-68)\right)} = 54.61°C$$

(From CH485; LMTD comes from integration of the stream enthalpy difference along length of exchanger.)

Multiple Passes

Use correction factor “F” described in FEE manual on page 214
Manual does not explain how to calculate it. Calculated in CHEMCAD.



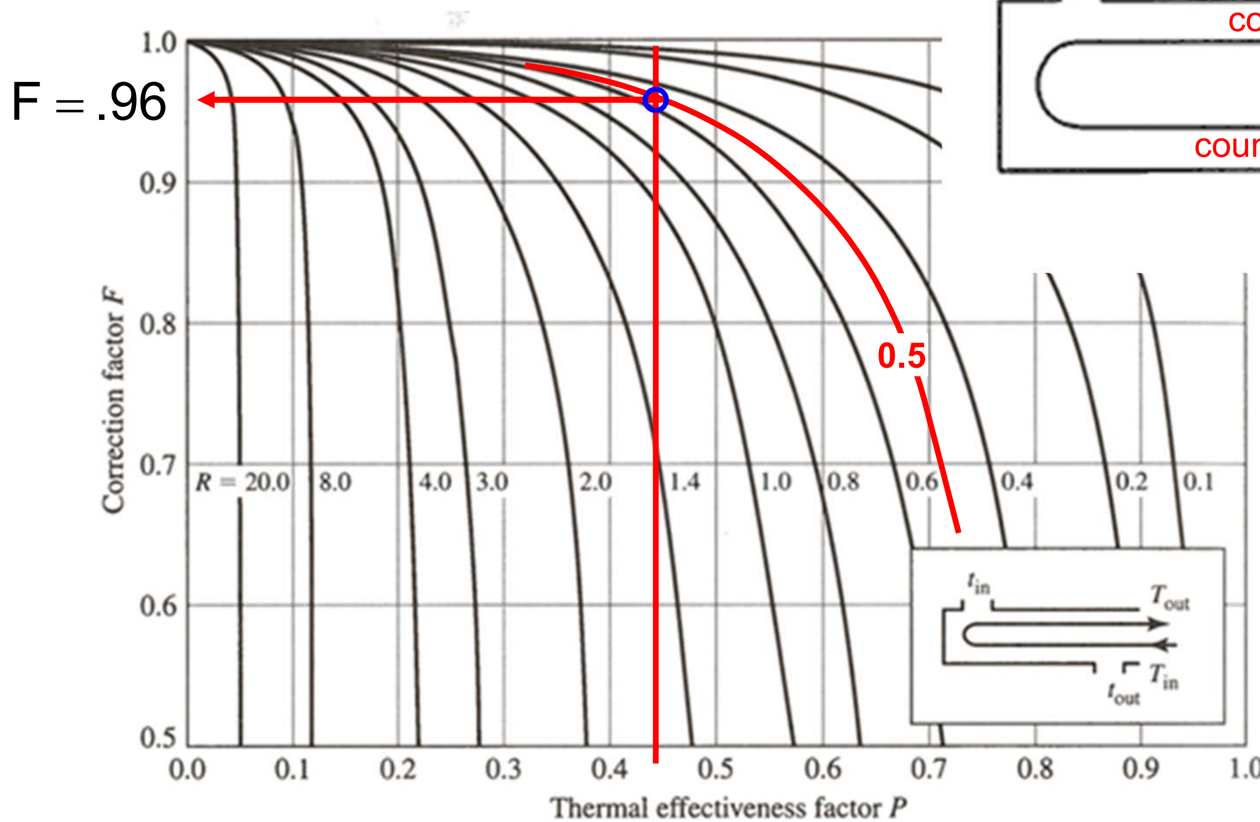


Figure 14-4. One shell pass and 2 or more tube passes.
Figure 14-5. Other contact patterns.

$$R = \frac{T_{Hi} - T_{Ho}}{T_{Co} - T_{Ci}} = \frac{158 - 138}{108 - 68} = 0.5$$

(14-9a, p.648)

$$P = \frac{T_{Co} - T_{Ci}}{T_{Hi} - T_{Ci}} = \frac{108 - 68}{158 - 68} = 0.44$$

(14-9b)

$$F = \frac{(R^2+1)^{1/2} \cdot \ln\left(\frac{1-P}{1-R\cdot P}\right)}{(R-1)\cdot \ln\left(\frac{2-P\cdot\left[(R+1)-(R^2+1)^{1/2}\right]}{2-P\cdot\left[(R+1)+(R^2+1)^{1/2}\right]}\right)} = 0.96$$

(14-9)

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

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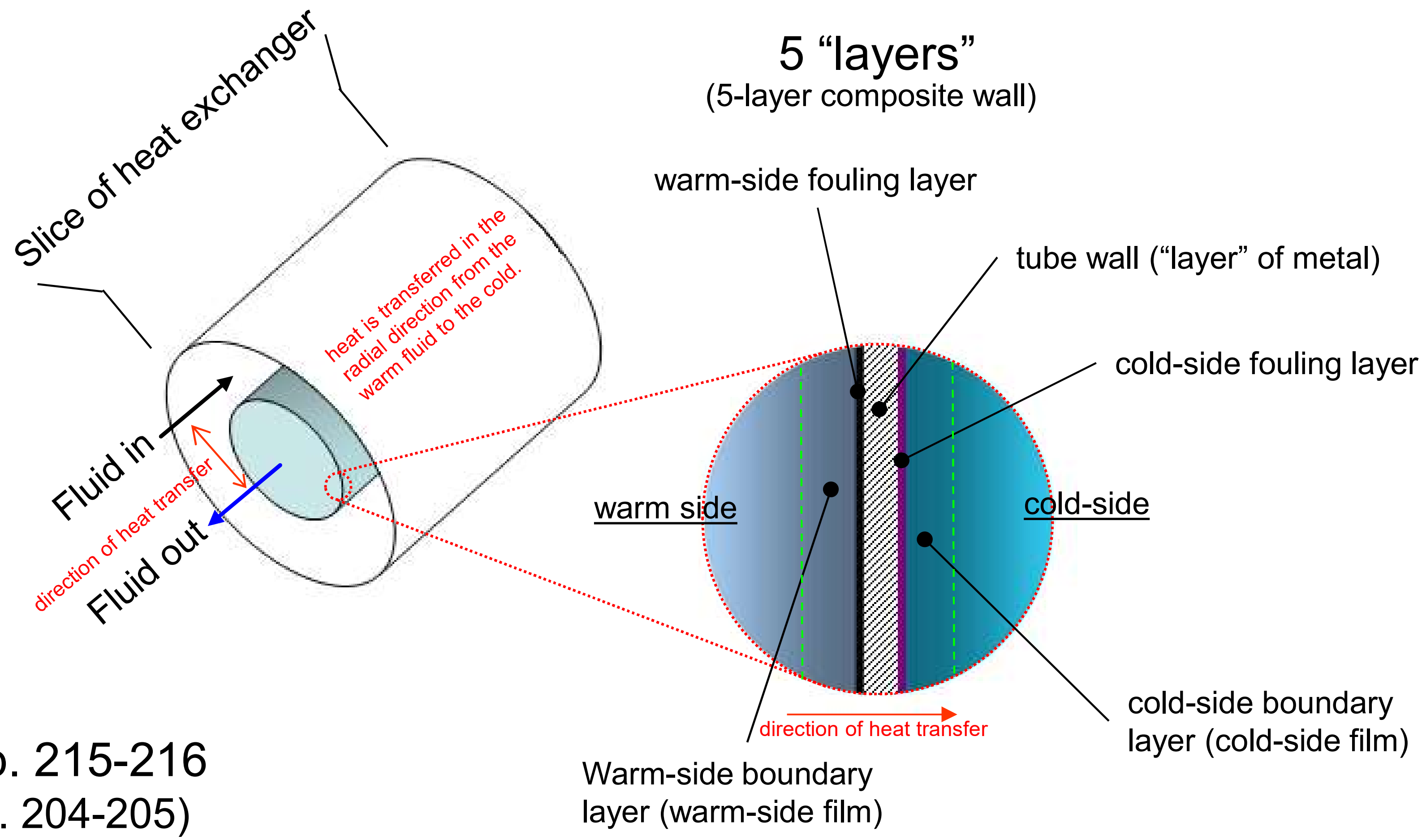
- pressure drop (slides 19-21)

Thermal Resistances in Series Model



FEE Manual, pp. 215-216
(explained on pp. 204-205)

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

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

FEE manual, page 216.

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

$A = A_i$ or A_o

$A_i = \pi \cdot D_i \cdot L$ (inside area)

$A_o = \pi \cdot D_o \cdot L$ (outside area)

L = length

FE Reference Manual, page 216 (222/502)

Example:

$A = A_o$

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

Local Heat Transfer Coefficients for Flow Inside Tubes

FE Prep: Heat Transfer Section, FEE Manual, p. 211

Needed for problem 14-2

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

FEE Manual, page 209

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

Turbulent, fully developed fluid flow inside tubes, Re>10,000, 0.7<Pr<16,7000, and L/D_i>10

14-18, p. 657

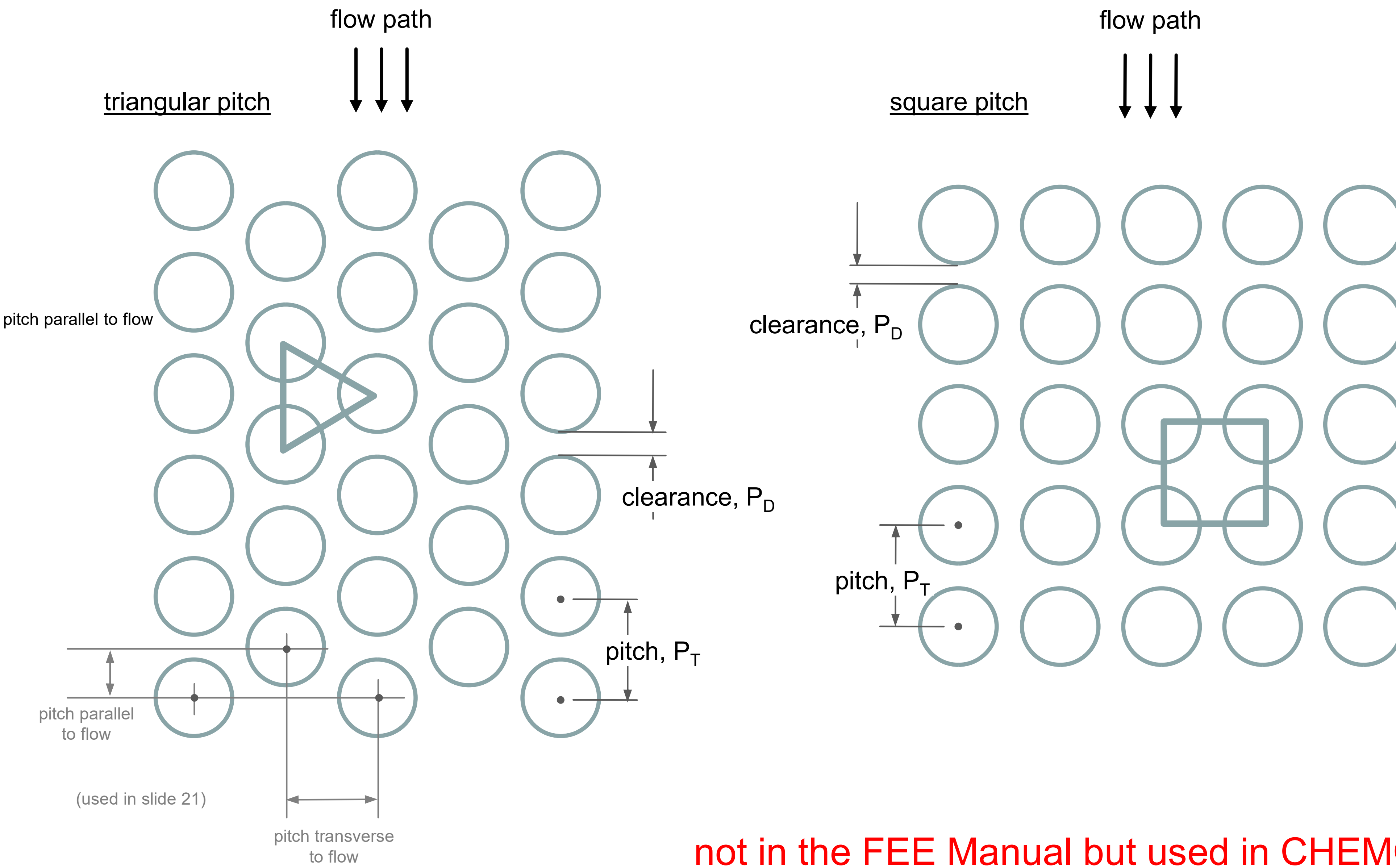
book typo here

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

Sieder-Tate equation
FEE Manual, page 210

(special case of Dittus-Boelter equation from CH485; viscosity term is dropped)

Tube Bank Patterns in Shell-and-Tube Heat Exchangers



not in the FEE Manual but used in CHEMCAD

Film Coefficients for Flow Outside 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

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

Eq. 14-22c, p. 661

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

Eq. 14-22a , p. 660

Constants a and m are in Table 14-1 and depend on shell-side Reynolds number:

Table 14-1 is in the FEE Manual on page 209

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

Calculate F_1:

FEE Manual method does not include F_1 or F_2 corrections.

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

Eq. 14-22b, p. 660

Look up F_2 Table 14-2

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

FEE Manual page 209

from Perry's Handbook

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

U = Btu/(°F · ft^2 · 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 · ft^2 · °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

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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

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

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

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

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

$S_i / S_H =$ 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·m²

Pressure Drop – Shell-side

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

Eq 14-24, p. 665

B_0 = number of tube crossings

(correction factor for re-crossing of tubes; B_0 = 1 for crossing of unbaffled tubes)

N_{tr} = number of tube rows

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

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

x_T is the ratio of pitch (P_T) transverse to flow to outside tube diameter

x_L is the ratio of pitch (P_T) parallel to flow to outside tube diameter

FEE tip:

G is known as the mass velocity and is equal to v·ρ where v is velocity and ρ is density.

$\left(\dfrac{D \cdot G}{\mu} \right)$ is another way to calculate Reynolds number

$$\left(\frac{D \cdot v \cdot \rho}{\mu} \right)$$

$$G = v \cdot \rho = \frac{m}{s} \cdot \frac{kg}{m^3} = \frac{kg}{s \cdot m^2}$$

Questions