LAMPIRAN

Shell (Residu)	Tube (Crude Oil)
Data Pengamatan : T1 = 447,8 ˚F T2 = 298,04 ˚F V = 71,97 ft^3/jam ρ = 0,9125 Kg/L	Data Pengamatan : t1 = 196,88 ˚F t2 = 269,6 ˚F V = 389,4 ft^3/jam ρ = 0,8381 Kg/L

Shell (Residu)	Tube (Crude Oil)
1. Menghitung Specific Gravity (Sg)	**1. Menghitung Specific Gravity (Sg)**
Diketahui : ρ = 0,9125 Kg/L	Diketahui : ρ = 0,8381 Kg/L

Shell (Residu):

Dens 15	Sg
0,913	0,9135
0,9125	0,9129
0,9124	0,9129

(ASTM Petroleum Measuresment)

Tube (Crude Oil):

Dens 15	Sg
0,8383	0,8388
0,8381	0,8385
0,8379	0,8383

(ASTM Petroleum Measuresment)

Shell (Residu)	Tube (Crude Oil)
2. Menghitung ˚API	**2. Menghitung ˚API**
Diketahui : ρ = 0,9125 Kg/L	Diketahui : ρ = 0,8381 Kg/L

Shell (Residu):

Dens 15	API
0,913	23,4
0,9125	23,483
0,9124	23,5

(ASTM Petroleum Measuresment)

Tube (Crude Oil):

Dens 15	API
0,8383	37,2
0,8381	37,3
0,8378	37,4

(ASTM Petroleum Measuresment)

Shell (Residu)	Tube (Crude Oil)
3. Menghitung Temperature Average (Tav)	**3. Menghitung Temperature Average (tav)**

$Tav = \frac{T1 + T2}{2}$ $Tav = \frac{447{,}8\ °F + 298{,}04°F}{2}$ $Tav = \frac{745{,}84\ °F}{2}$ $Tav = 372{,}92\ °F$	$tav = \frac{T1 + T2}{2}$ $tav = \frac{196{,}88\ °F + 269{,}6\ °F}{2}$ $tav = \frac{466{,}48\ °F}{2}$ $tav = 233{,}24\ °F$
4. Menghitung Cp	**4. Menghitung Cp**
Dari Fig. 4 Kern Tav = 372,92 ˚F, maka Cp = 0,59 Btu/lb ˚F	Dari Fig. 4 Kern Tav = 233,24 ˚F, maka Cp = 0,55 Btu/lb ˚F
5. Menghitung Massa Jenis	**5. Menghitung Massa Jenis**
ρ = 0,9125 Kg/L ρ = 56,97 lb/ft^3	ρ = 0,8381 Kg/L ρ = 52,322 lb/ft^3
6. Menghitung Mass Flow	**6. Menghitung Mass Flow**
W Residu = ρ Residu.V Residu W Residu = 56,97 lb/ft^3 x 71,97 ft^3/jam W Residu = 4067,5 lb/jam	W Co = ρ Co.V Co W Co = 52,322 lb/ft^3 x 389,4 ft^3/jam W Co = 20374 lb/jam
7. Menghitung nilai Panas yang dibutuhkan (Q)	**7. Menghitung nilai Panas yang dibutuhkan (Q)**
ΔT Residu = T1 – T2 ΔT Residu = 447,8 ˚F - 447,8 ˚F ΔT Residu = 149,76 ˚F Q Residu = W Residu × Cp Residu × ΔT Residu Q Residu = 4067,5 lb/jam x 0,59 Btu/lb ˚F x 149,76 ˚F Q Residu = 359397 Btu/jam	ΔT Crude Oil = t2 – t1 ΔT Crude Oil = 269,6 ˚F - 196,88 ˚F ΔT Crude Oil = 72,72 ˚F Q Co = W Co × Cp Co × ΔT Co Q Co = 20374 lb/jam x Q Co = 359397 Btu/jam x 0,55 Btu/lb ˚F x 72,72 ˚F Q Co = 814886 Btu/jam
8. Menghitung Neraca Panas	
Q = Q Co – Q Residu	

Q = 814886 Btu/jam - 359397 Btu/jam

Q = 455489,324 Btu/jam

Q Losses = (Q/Q Co) x 100%

Q Losses = (455489,324 Btu/jam/814886 Btu/jam) x 100%

Q Losses = 55,896 %

9. Menghitung LMTD

Hot Fluid (F)		Cold Fluid (F)	Diff	
447,8	High T	269,6	178,2	Δt2
298,04	Low T	196,88	101,16	Δt1
			77,04	Δt2-Δt1

$$LMTD = \frac{\Delta t2 - \Delta t1}{\ln\left(\frac{\Delta t2}{\Delta t1}\right)}$$

$$LMTD = \frac{178{,}2\ ^{\circ}\mathrm{F} - 101{,}16\ ^{\circ}\mathrm{F}}{ln\,\frac{178{,}2\ ^{\circ}\mathrm{F}}{101{,}16\ ^{\circ}\mathrm{F}}}$$

$$LMTD = 136{,}06\ ^{\circ}\mathrm{F}$$

$$R = \frac{T1 - T2}{t2 - t1}$$

$$R = \frac{447{,}8\ ^{\circ}\mathrm{F} - 298{,}04\ ^{\circ}\mathrm{F}}{269{,}6\ ^{\circ}\mathrm{F} - 196{,}88\ ^{\circ}\mathrm{F}}$$

$$R = 2{,}0594$$

$$S = \frac{t2 - t1}{T1 - t1}$$

$$S = \frac{269{,}6\ ^{\circ}\mathrm{F} - 196{,}88\ ^{\circ}\mathrm{F}}{447{,}8\ ^{\circ}\mathrm{F} - 196{,}88\ ^{\circ}\mathrm{F}}$$

$$S = 0{,}2898$$

Diperoleh nilai Ft = 0,9 (Fig.18 Kern)

ΔTLMTD = LMTD x Ft

ΔTLMTD = $136{,}06\ ^{\circ}\mathrm{F} \times 0{,}9$

ΔTLMTD = 122,46 ˚F

10. Menentukan Caloric Temperature

a. Menentukan nilai Kc

ΔT Residu = T1 – T2

ΔT Residu = 447,8 ˚F - 298,04 ˚F

ΔT Residu = 149,76 ˚F

˚API = 23,483

Kc = 0,65 (Fig.17 Kern)

b. ΔT Cold/ ΔT Hot

ΔT Cold/ ΔT Hot = 0,567

Fc = 0,41 (Fig.17 Kern)

c. Menentukan Calaoric Temperature

Tc Residu = T2 + Fc (T1 – T2)

Tc Residu = 298,04 ˚F + 0,41 (447,8 ˚F-298,04 ˚F)

Tc Residu = 359,44 ˚F

tc crude oil = t1 + Fc x (t2 – t1)

tc crude oil = 196,88 ˚F + 0,41 (269,6 ˚F- 196,88 ˚F)

tc crude oil = 226,7 ˚F

11. Menentukan Flow Area	11. Menentukan Flow Area
ID Shell = 36,457 i C' = 0,25 i B = 25,886 in n = 1 Pt = 1,25 in $as = \frac{ID\ shell \times C' \times B}{n \times 144 \frac{in^2}{ft^2} \times Pt}$	Nt = 400 n = 1 ODtube = 1 BWG = 14 a' = 0,479 (Kern, Tabel 10) $at = \frac{Nt \times a'}{144 \frac{in^2}{ft^2} \times n}$

$$as = \frac{36{,}457 \text{ in } \times \text{ } 0{,}25 \text{ in } \times \text{ } 25{,}886 \text{ in}}{1 \times 144 \frac{in^2}{ft^2} \times 1{,}25 \; in}$$

$$as = 1{,}3107 \; ft^2$$

$$at = \frac{400 \times 0{,}479}{144 \frac{in^2}{ft^2} \times 1}$$

$$at = 1{,}3305 \; ft^2$$

12. Menentukan Mass Velocity

$$Gs = \frac{W \; Residu}{as}$$

$$Gs = \frac{4067{,}5 \frac{lb}{jam}}{1{,}3107 \; ft^2}$$

$$Gs = 3103{,}2 \; \frac{lb}{jam \; ft^2}$$

12. Menentukan Mass Velocity

$$Gt = \frac{W \; Crude \; Oil}{at}$$

$$Gs = \frac{20374{,}2 \frac{lb}{jam}}{1{,}3305 \; ft^2}$$

$$Gs = 15312{,}5 \; \frac{lb}{jam \; ft^2}$$

13. Menentukan Bilangan Reynolds

Diketahui :

OD = 1 in

Pt = 1,25

De = 0,72 in

= 0,0598 ft (Fig.28 Kern)

˚API = 23,483

API	X	Y
34	10,3	21,3
28	10	23,6
23,483	9,977	25,331

Tc = 359,44 ˚F

μ = 1,2 Cp

= 2,904 lb/ft jam

$$Re = \frac{De \; x \; Gs}{\mu}$$

$$Re = \frac{0{,}0598 \text{ ft x } 3103{,}2 \; \frac{lb}{jam \; ft^2}}{2{,}904 \text{ lb/ft jam}}$$

13. Menentukan Bilangan Reynolds

Diketahui :

D Tube = 0,782 in (Tabel 10 Kern)

= 0,06514 ft

˚API = 37,3

API	X	Y
42	11,6	16
37,3	10,52	18,6857
35	10	20

tc = 226,695 ˚F

μ = 0,8 Cp

= 1,936 lb/ft jam

$$Re = \frac{De \; x \; Gs}{\mu}$$

$Re = 63{,}86$	$Re = \dfrac{0{,}06514\ ft \times 15312{,}5\ \frac{lb}{jam\ ft^2}}{1{,}936\ \frac{lb}{ft} jam}$ $Re = 515{,}221$
14. Menentukan Faktor Perpindahan Panas	**14. Menentukan Faktor Perpindahan Panas**
Jhs = 3,9 (Fig.28 Kern)	Diketahui : L = 11,482 ft ID = 0,06514 ft L/ID = 11,482 ft/0,06514 ft Jht = 2,5 (Fig.24 Kern)
15. Menentukan Koefisien Perpindahan Panas	**15. Menentukan Koefisien Perpindahan Panas**
Diketahui : Tc = 359,44 ˚F ˚API = 23,483 K = 0,067 (Fig. 1 Kern) $ho = jHs \times \frac{k}{De} \times \left(\frac{Cp \times \mu}{k}\right)^{1/3} \times \left(\frac{\mu}{\mu_w}\right)^{0,14}$ $\frac{ho}{\emptyset s} = 3{,}0 \times \frac{0{,}067}{0{,}0598} \times \left(\frac{0{,}59 \times 1{,}2}{0{,}067}\right)^{1/3}$ $\frac{ho}{\emptyset s} = 11{,}563\ \frac{Btu}{jam\ ft^2\ °F}$	Diketahui : tc = 226,7 ˚F K = 0,079 (Fig.1 Kern) $hi = jHt \times \frac{k}{De} \times \left(\frac{Cp \times \mu}{k}\right)^{\frac{1}{3}} \times \left(\frac{\mu}{\mu_w}\right)^{0,14}$ $\frac{ho}{\emptyset t} = 2{,}5 \times \frac{0{,}079}{0{,}06514} \times \left(\frac{0{,}55 \times 0{,}8}{0{,}079}\right)^{1/3}$ $\frac{hi}{\emptyset t} = 6{,}61618 \frac{Btu}{jam.\ ft^2.\ °F}$ Diketahui : OD Tube = 1 in ID = 0,782 in (Tabel 10 Kern) $\frac{h_{io}}{\emptyset t} = \frac{hi}{\emptyset t} \times \frac{Di}{OD}$ $\frac{h_{io}}{\emptyset t} = 6{,}61618 \frac{Btu}{hr.\ ft^2.\ °F} \times \frac{0{,}782\ in}{1\ in}$ $\frac{h_{io}}{\emptyset t} = 5{,}1738 \frac{Btu}{hr.\ ft^2.\ °F}$
16. Menentukan Wall Temperature	

$$tw = tc + \frac{\frac{ho}{\emptyset s}}{\frac{hio}{\emptyset t} + \frac{ho}{\emptyset s}}(Tc - tc)$$

$$tw = 226{,}7\ °F + \frac{11{,}563\ \frac{Btu}{jam.ft^2.°F}}{5{,}1738\frac{Btu}{jam.ft^2.°F} + 11{,}563\ \frac{Btu}{jam.ft^2.°F}}(359{,}44\ °F - 226{,}7\ °F\ )$$

$$tw = 318{,}41\ °F$$

17. Menentukan Koefisien Transfer Panas

Diketahui :

tw = 318,41 ˚F

˚API = 23,483

API	X	Y
34	10	20
28	10,8	21,3
23,483	11,4	19,021

μ = 0,5 Cp

= 1,21 lb/ft jam

$$\emptyset_s = \left(\frac{\mu}{\mu_w}\right)^{0,14}$$

$$\emptyset_s = \left(\frac{2{,}904}{1{,}21}\right)^{0,14}$$

$$\emptyset_s = 1{,}1304$$

17. Menentukan Koefisien Transfer Panas

Diketahui :

tw = 318,41 ˚F

˚API = 37,3

API	X	Y
42	11,6	16
37,3	10,52	18,68
35	10	20

μ = 0,35 Cp

= 0,847 lb/ft jam

$$\emptyset_s = \left(\frac{\mu}{\mu_w}\right)^{0,14}$$

$$\emptyset_s = \left(\frac{1{,}936}{0{,}847}\right)^{0,14}$$

$$\emptyset_s = 1{,}1227$$

18. Menentukan Koefisien Perpindahan Panas Terkoreksi

$$ho = \frac{ho}{\emptyset s} \times \emptyset s$$

$$ho = 11{,}563\ \frac{Btu}{hr.ft^2.°F} \times 1{,}1304$$

18. Menentukan Koefisien Perpindahan Panas Terkoreksi

$$hi = \frac{hi}{\emptyset t} \times \emptyset t$$

$$hi = 5{,}1738\frac{Btu}{hr.ft^2.°F} \times 1{,}1227$$

$ho = 13{,}07 \frac{Btu}{hr.ft^2.°F}$	$hi = 5{,}8086 \frac{Btu}{hr.ft^2.°F}$
19. Menentukan Clear Overall Coefficient	
$Uc = \frac{ho \times hio}{ho + hio}$ $Uc = \frac{13{,}07 \frac{Btu}{hr.ft^2.°F} \times 5{,}8086 \frac{Btu}{hr.ft^2.°F}}{13{,}07 \frac{Btu}{hr.ft^2.°F} + 5{,}8086 \frac{Btu}{hr.ft^2.°F}}$ $Uc = 5{,}5343 \frac{Btu}{hr.ft^2.°F}$	
20 Menentukan Fouling Factor	
Diketahui : Nt = 400 OD = 1 BWG = 14, maka didapatkan a''= 0,2618 (Tabel 10 Kern) L = 11,483 ft $A = Nt \times a'' \times L$ $A = 400 \times 0{,}2618 \frac{ft^2}{ft} \times 11{,}482\, ft$ $A = 1202{,}39504\, ft^2$ $UD = \frac{Qt}{A \times \Delta TLMTD}$ $Ud = \frac{814886 \frac{Btu}{jam}}{1202{,}39504\, ft^2 \times 122{,}46\, °F}$ $Ud = 4{,}0215 \frac{Btu}{hr.ft^2.°F}$ $Rd = \frac{UC - UD}{UC \times UD}$ $Rd = \frac{5{,}5343 \frac{Btu}{hr.ft^2.°F} - 4{,}0215 \frac{Btu}{hr.ft^2.°F}}{5{,}5343 \frac{Btu}{hr.ft^2.°F} \times 4{,}0215 \frac{Btu}{hr.ft^2.°F}}$ $Rd = 0{,}068$	

21. Menentukan Effisiensi	
$\eta = \frac{Q\ Shell}{Q\ Tube} \times 100\%$ $\eta = \frac{359397 \frac{Btu}{hr}}{814886 \frac{Btu}{hr}} \times 100\%$ $\eta = 44{,}104\ \%$	
22. Evaluasi Pressure Drop	
1. Menentukan Nilai Friksi	**1. Menentukan Nilai Friksi**
Diketahui : Res = 63,86 f = 0,008 (Fig.29 Kern)	Diketahui : Ret = 515,221 f = 0,00032 (Fig.26 Kern)
2. Pressure Drop	
Diketahui : L = 11,482 ft B = 25,866 in $N + 1 = \frac{12 \times L}{B}$ $N + 1 = \frac{12 \times 11{,}482\ ft}{25{,}866\ in}$ $N + 1 = 5{,}326\ ft$ De = 0,0598 ft ID = 3,0259 ft f = 0,008 Gs = 3103,2 lb/jam ft^2 Sg = 0,9129 $\emptyset_s$ = 1,1304 $\Delta Ps = \frac{f \times Gs^2 \times IDs \times (N+1)}{5{,}22 \times 10^{10} \times De \times SG \times \emptyset s}$ $\Delta Ps = \frac{0{,}008 \times (3103{,}2)^2 \times 3{,}0259 \times 5{,}326}{5{,}22 \times 10^{10} \times 0{,}0598 \times 0{,}9129 \times 1{,}1304}$	L = 11,482 ft n = 1 ID = 0,0649 ft Gt = 15312,5412 lb/jam ft^2 Sg = 0,8385 $\emptyset_s$ = 1,1226 f = 0,00032 $\Delta Pt = \frac{f \times Gt^2 \times L \times n}{5{,}22 \times 10^{10} \times Di \times SG \times \emptyset t}$ $\Delta Pt = \frac{0{,}00032 \times (15312{,}5412)^2 \times 11{,}482 \times 1}{5{,}22 \times 10^{10} \times 0{,}0649 \times 0{,}8385 \times 1{,}1226}$ $\Delta Pt = 0{,}00027\ Psi$

$\Delta Ps = 0{,}00039\ Psi$	

3. Menentukan ΔP Total

ρ (Crude Oil) = 52,322 lb/ft^3

Gt = 15313 lb/jam ft^2

n = 1

Sg = 0,8385

g = 32,2 ft^2/s

$$v = \frac{Gt}{\rho}$$

$$v = \frac{15313\,\frac{lb}{jam\ ft^2}}{52{,}322\ \frac{lb}{ft^3}}$$

$$v = 292{,}66\ ft/jam$$

$$\Delta Pr = \frac{4n}{SG} \times \frac{v}{2g}$$

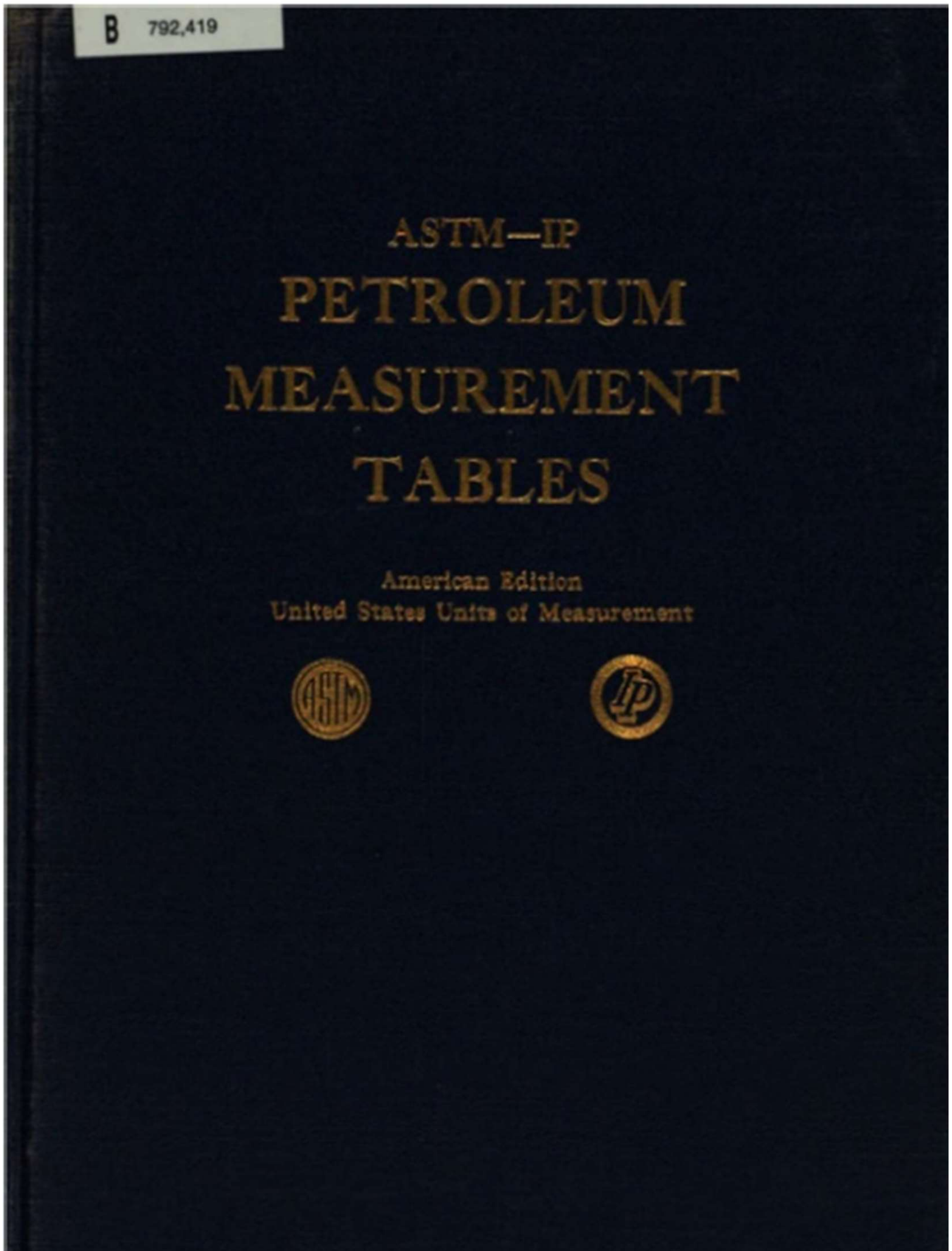
$$\Delta Pr = \frac{4 \times 1}{0{,}8385} \times \frac{292{,}66\,\frac{ft}{jam}}{2 \times 32{,}2\,\frac{ft^2}{s}}$$

$$\Delta Pr = 0{,}00048\ Psi$$

ΔP total = ΔPt + ΔPr

ΔP total = 0,00027 psi + 0,00048 psi

ΔP total = 0,00075 psi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ASTM—IP
PETROLEUM
MEASUREMENT
TABLES
American Edition
United States Units of Measurement

Table 3

ASTM—IP **API Gravity to Specific Gravity and to Density** **15–30° API**

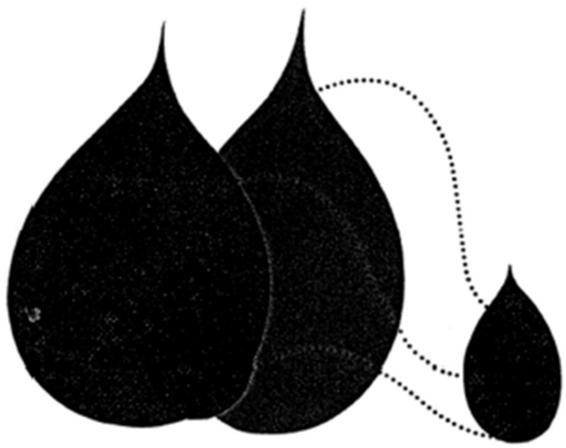
API Gravity 60°F.	Specific Gravity 60/60°F.	Density 15°C.	API Gravity 60°F.	Specific Gravity 60/60°F.	Density 15°C.	API Gravity 60°F.	Specific Gravity 60/60°F.	Density 15°C.
15.0	0.9659	0.9653	20.0	0.9340	0.9335	25.0	0.9042	0.9037
15.1	0.9652	0.9647	20.1	0.9334	0.9328	25.1	0.9036	0.9031
15.2	0.9646	0.9640	20.2	0.9328	0.9322	25.2	0.9030	0.9025
15.3	0.9639	0.9633	20.3	0.9321	0.9316	25.3	0.9024	0.9019
15.4	0.9632	0.9627	20.4	0.9315	0.9310	25.4	0.9018	0.9014
15.5	0.9626	0.9620	20.5	0.9309	0.9304	25.5	0.9013	0.9008
15.6	0.9619	0.9614	20.6	0.9303	0.9298	25.6	0.9007	0.9002
15.7	0.9613	0.9607	20.7	0.9297	0.9292	25.7	0.9001	0.8996
15.8	0.9606	0.9601	20.8	0.9291	0.9286	25.8	0.8996	0.8991
15.9	0.9600	0.9594	20.9	0.9285	0.9280	25.9	0.8990	0.8985
16.0	0.9593	0.9588	21.0	0.9279	0.9273	26.0	0.8984	0.8979
16.1	0.9587	0.9581	21.1	0.9273	0.9267	26.1	0.8978	0.8974
16.2	0.9580	0.9575	21.2	0.9267	0.9261	26.2	0.8973	0.8968
16.3	0.9574	0.9568	21.3	0.9260	0.9255	26.3	0.8967	0.8962
16.4	0.9567	0.9562	21.4	0.9254	0.9249	26.4	0.8961	0.8956
16.5	0.9561	0.9555	21.5	0.9248	0.9243	26.5	0.8956	0.8951
16.6	0.9554	0.9549	21.6	0.9242	0.9237	26.6	0.8950	0.8945
16.7	0.9548	0.9542	21.7	0.9236	0.9231	26.7	0.8944	0.8940
16.8	0.9541	0.9536	21.8	0.9230	0.9225	26.8	0.8939	0.8934
16.9	0.9535	0.9530	21.9	0.9224	0.9219	26.9	0.8933	0.8928
17.0	0.9529	0.9523	22.0	0.9218	0.9213	27.0	0.8927	0.8923
17.1	0.9522	0.9517	22.1	0.9212	0.9207	27.1	0.8922	0.8917
17.2	0.9516	0.9510	22.2	0.9206	0.9201	27.2	0.8916	0.8911
17.3	0.9509	0.9504	22.3	0.9200	0.9195	27.3	0.8911	0.8906
17.4	0.9503	0.9498	22.4	0.9194	0.9189	27.4	0.8905	0.8900
17.5	0.9497	0.9491	22.5	0.9188	0.9183	27.5	0.8899	0.8895
17.6	0.9490	0.9485	22.6	0.9182	0.9177	27.6	0.8894	0.8889
17.7	0.9484	0.9478	22.7	0.9176	0.9171	27.7	0.8888	0.8883
17.8	0.9478	0.9472	22.8	0.9170	0.9165	27.8	0.8883	0.8878
17.9	0.9471	0.9466	22.9	0.9165	0.9159	27.9	0.8877	0.8872
18.0	0.9465	0.9459	23.0	0.9159	0.9153	28.0	0.8871	0.8867
18.1	0.9459	0.9453	23.1	0.9153	0.9148	28.1	0.8866	0.8861
18.2	0.9452	0.9447	23.2	0.9147	0.9142	28.2	0.8860	0.8856
18.3	0.9446	0.9441	23.3	0.9141	0.9136	28.3	0.8855	0.8850
18.4	0.9440	0.9434	23.4	0.9135	0.9130	28.4	0.8849	0.8845
18.5	0.9433	0.9428	23.5	0.9129	0.9124	28.5	0.8844	0.8839
18.6	0.9427	0.9422	23.6	0.9123	0.9118	28.6	0.8838	0.8833
18.7	0.9421	0.9415	23.7	0.9117	0.9112	28.7	0.8833	0.8828
18.8	0.9415	0.9409	23.8	0.9111	0.9106	28.8	0.8827	0.8822
18.9	0.9408	0.9403	23.9	0.9106	0.9100	28.9	0.8822	0.8817
19.0	0.9402	0.9397	24.0	0.9100	0.9095	29.0	0.8816	0.8811
19.1	0.9396	0.9390	24.1	0.9094	0.9089	29.1	0.8811	0.8806
19.2	0.9390	0.9384	24.2	0.9088	0.9083	29.2	0.8805	0.8801
19.3	0.9383	0.9378	24.3	0.9082	0.9077	29.3	0.8800	0.8795
19.4	0.9377	0.9372	24.4	0.9076	0.9071	29.4	0.8794	0.8790
19.5	0.9371	0.9366	24.5	0.9071	0.9065	29.5	0.8789	0.8784
19.6	0.9365	0.9359	24.6	0.9065	0.9060	29.6	0.8783	0.8779
19.7	0.9358	0.9353	24.7	0.9059	0.9054	29.7	0.8778	0.8773
19.8	0.9352	0.9347	24.8	0.9053	0.9048	29.8	0.8772	0.8768
19.9	0.9346	0.9341	24.9	0.9047	0.9042	29.9	0.8767	0.8762
20.0	0.9340	0.9335	25.0	0.9042	0.9037	30.0	0.8762	0.8757

11

PROCESS HEAT TRANSFER

D. Q. KERN

INTERNATIONAL STUDENT EDITION



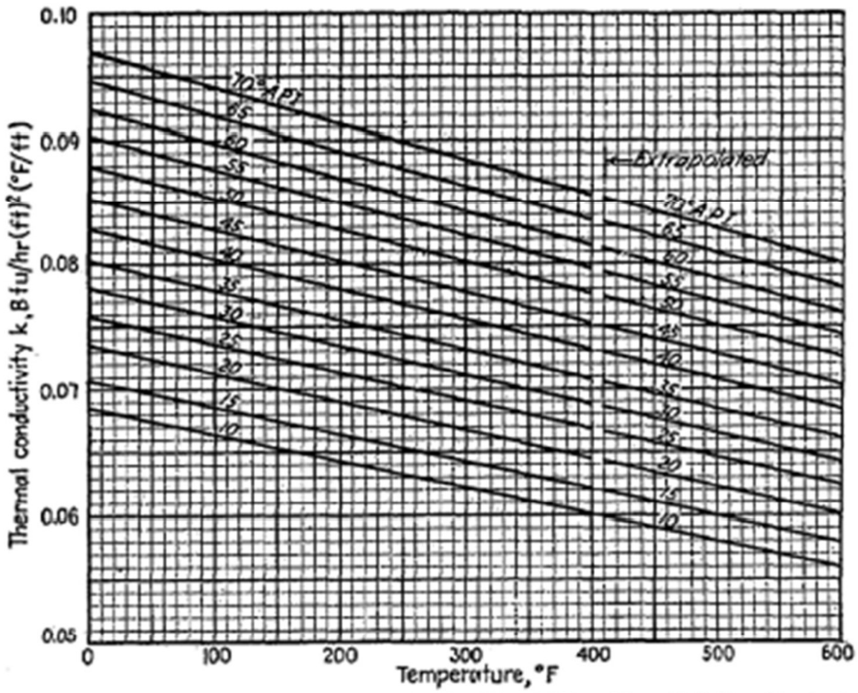


FIG. 1. Thermal conductivities of hydrocarbon liquids. (*Adapted from Natl. Bur. Standards Misc. Pub. 97.*)

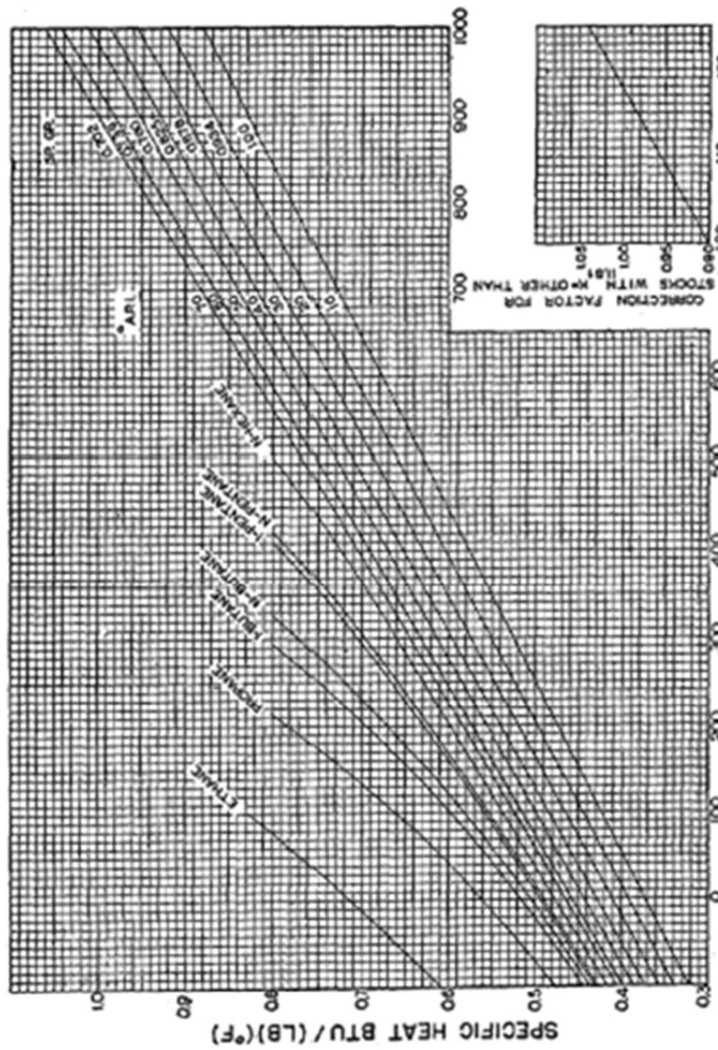


FIG. 4. Specific heats of hydrocarbon liquids. [*Holcomb and Brown, Ind. Eng. Chem.*, **34**, 595 (1942).]
† K = characterization factor.

VISCOSITIES OF PETROLEUM FRACTIONS
For temperature ranges employed in the text
Coordinates to be used with Fig. 14

	X	Y
76°API natural gasoline	14.4	6.4
56°API gasoline	14.0	10.5
42°API kerosene	11.6	16.0
35°API distillate	10.0	20.0
34°API mid-continent crude	10.3	21.3
28°API gas oil	10.0	23.6

VISCOSITIES OF ANIMAL AND VEGETABLE OILS*

	Acid No.	Sp gr, 20/4°C	X	Y
Almond	2.85	0.9188	6.9	28.2
Coconut	0.01	0.9226	6.9	26.9
Cod liver	.....	0.9138	7.7	27.7
Cottonseed	14.24	0.9187	7.0	28.0
Lard	3.39	0.9138	7.0	28.2
Linseed	3.42	0.9297	6.8	27.5
Mustard	.....	0.9237	7.0	28.5
Neatsfoot	13.35	0.9158	6.5	28.0
Olive	.....	0.9158	6.6	28.3
Palm kernel	9.0	0.9190	7.0	26.9
Perilla, raw	1.36	0.9297	8.1	27.2
Rapeseed	0.34	0.9114	7.0	28.8
Sardine	0.57	0.9384	7.7	27.3
Soybean	3.50	0.9228	8.3	27.5
Sperm	0.80	0.8829	7.7	26.3
Sunflower	2.76	0.9207	7.5	27.6
Whale, refined	0.73	0.9227	7.5	27.5

* Based on data at 100 and 210°F of A. R. Rescoria and F. L. Carnahan, *Ind. Eng. Chem.*, **28**, 1212–1213 (1936).

VISCOSITIES OF COMMERCIAL FATTY ACIDS*
250 to 400°F

	Sp gr at 300°F	X	Y
Lauric	0.792	10.1	23.1
Oleic	0.799	10.0	25.2
Palmitic	0.786	9.2	25.9
Stearic	0.789	10.5	25.5

* From data of D. Q. Kern and W. Van Nostrand, *Ind. Eng. Chem.*, **41**, 2209 (1949).

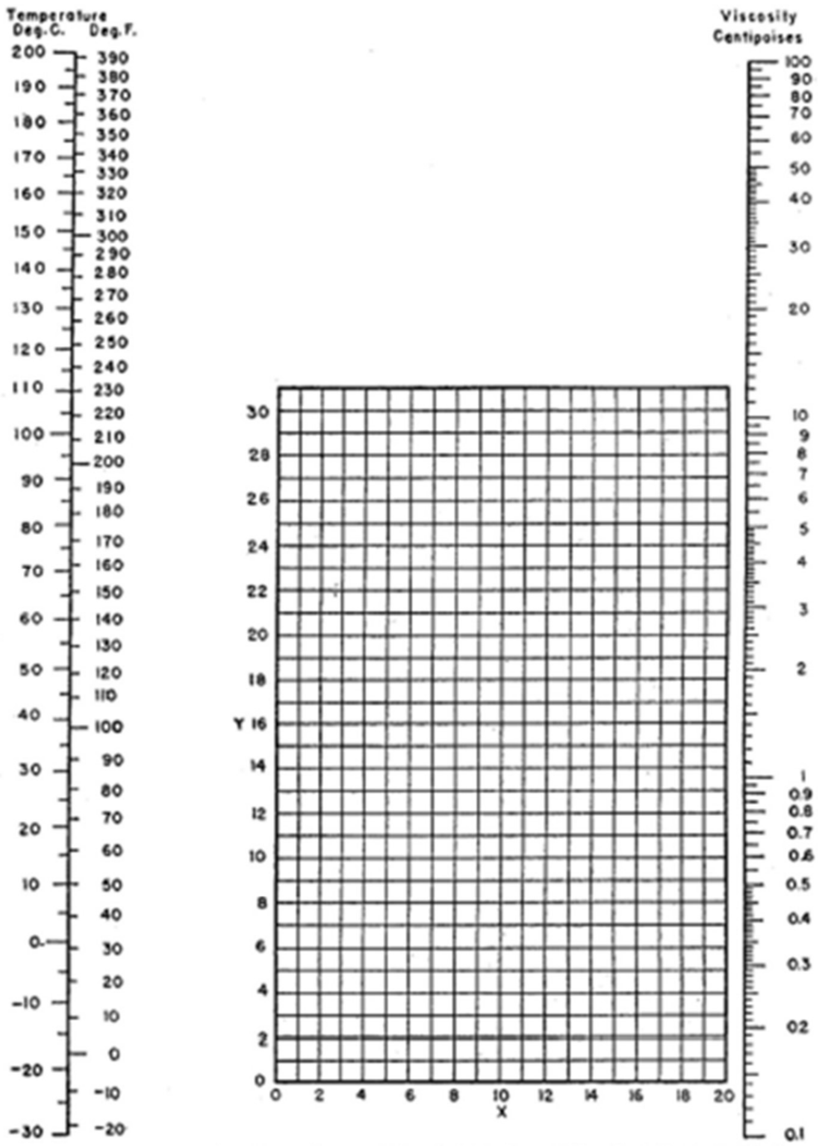


FIG. 14. Viscosities of liquids. (*Perry, "Chemical Engineers' Handbook," 3d ed., McGraw Hill Book Company, Inc., New York,* 1950.)

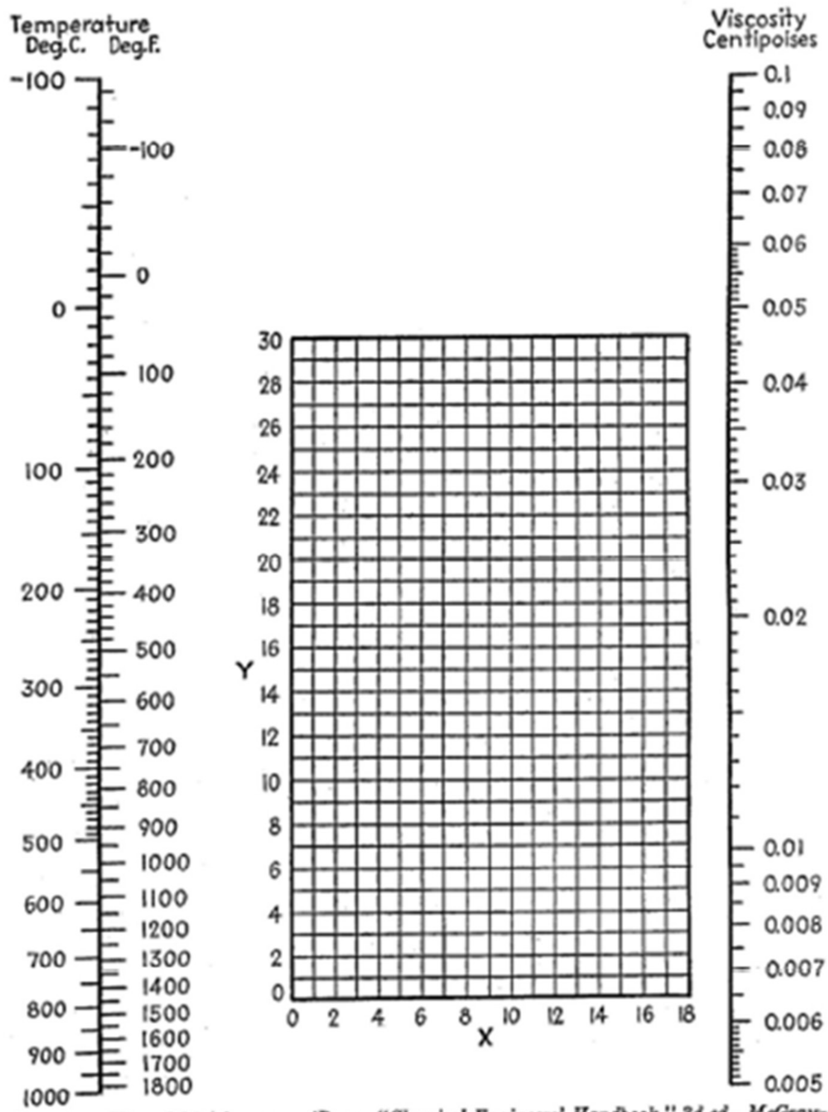


FIG. 15. Viscosities of gases. (*Perry, "Chemical Engineers' Handbook," 3d ed., McGraw-Hill Book Company, Inc., New York,* 1950.)

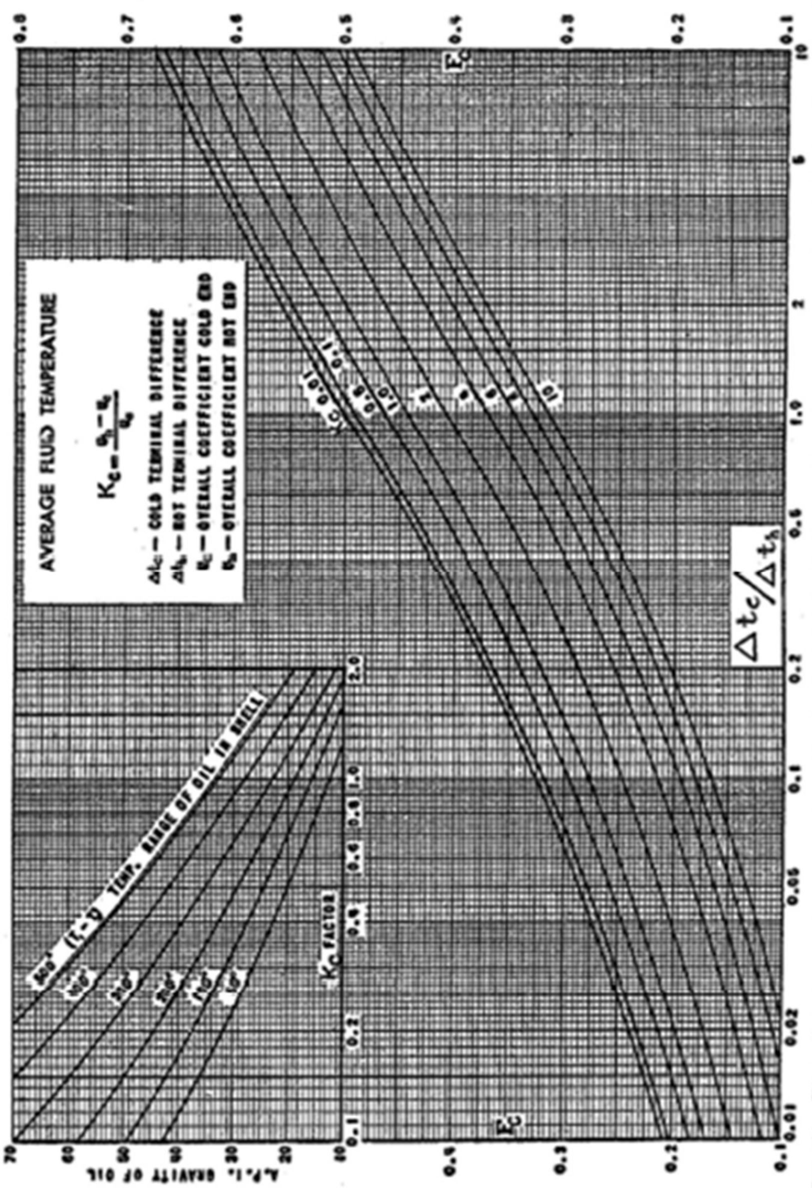


FIG. 17. The caloric temperature factor F_c. (*Standards of Tubular Exchanger Manufacturers Association, 2d ed., New York, 1949.*)

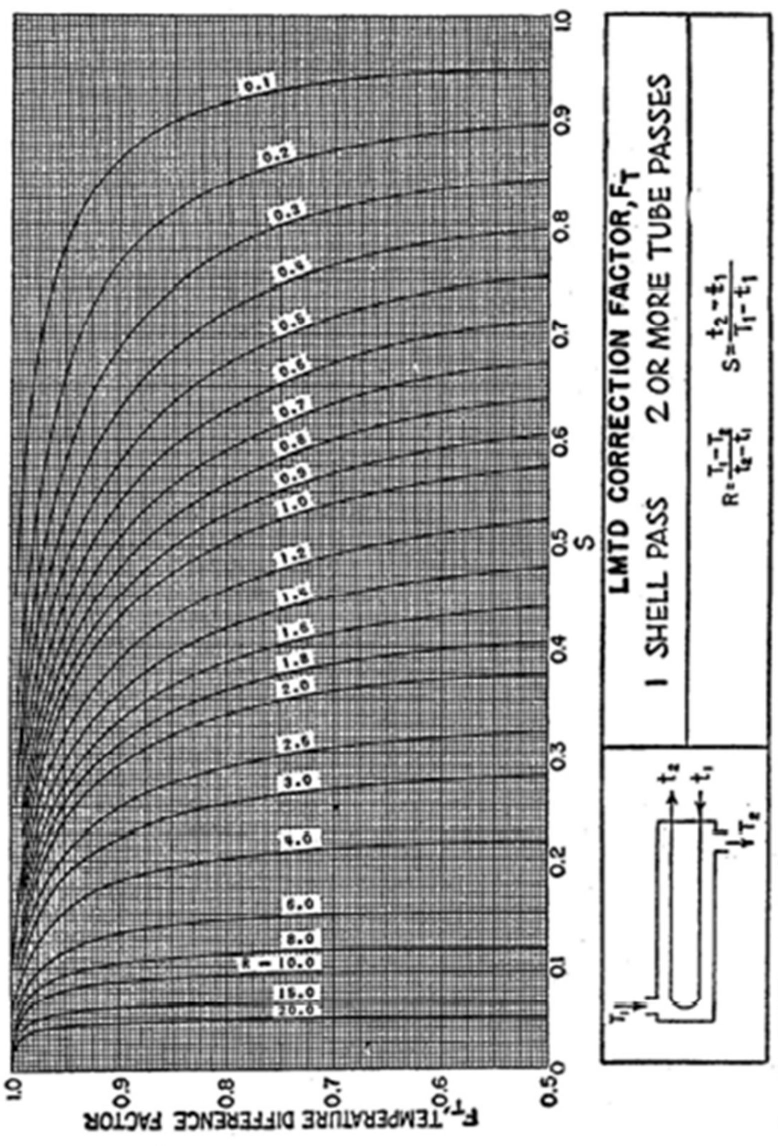


FIG. 18. LMTD correction factors for 1-2 exchangers. (*Standards of Tubular Exchanger Manufacturers Association*, 2d ed., New York, 1949.)

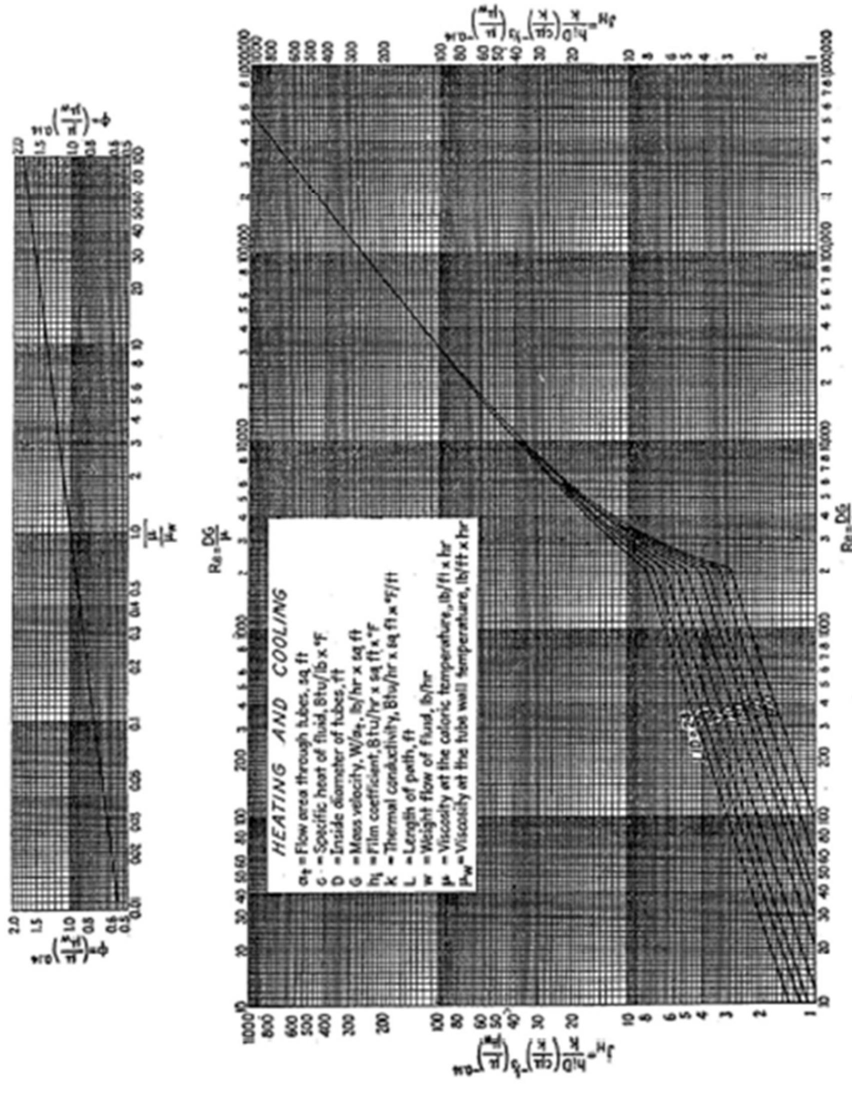


FIG. 24. Tube-side heat-transfer curve. (*Adapted from Sieder and Tate.*)

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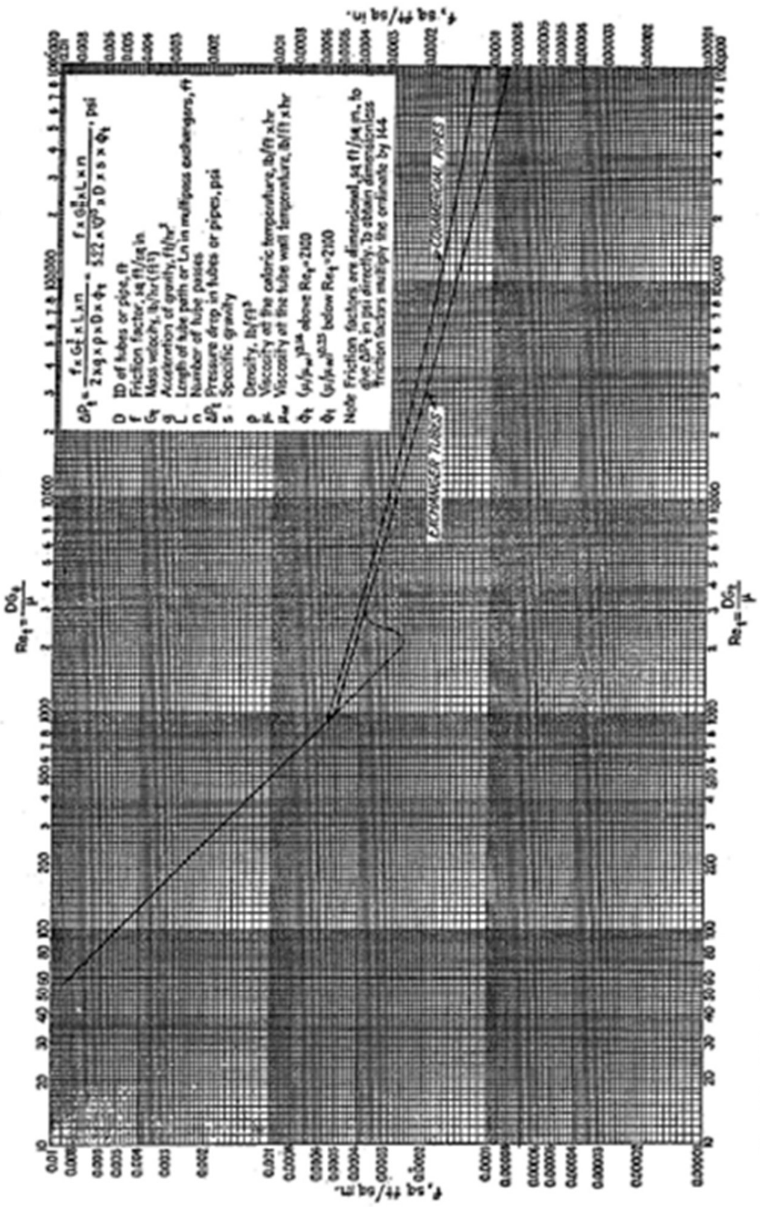

FIG. 26. Tube-side friction factors. (*Standards of Tubular Exchanger Manufacturers Association, 2d ed., New York, 1949.*)

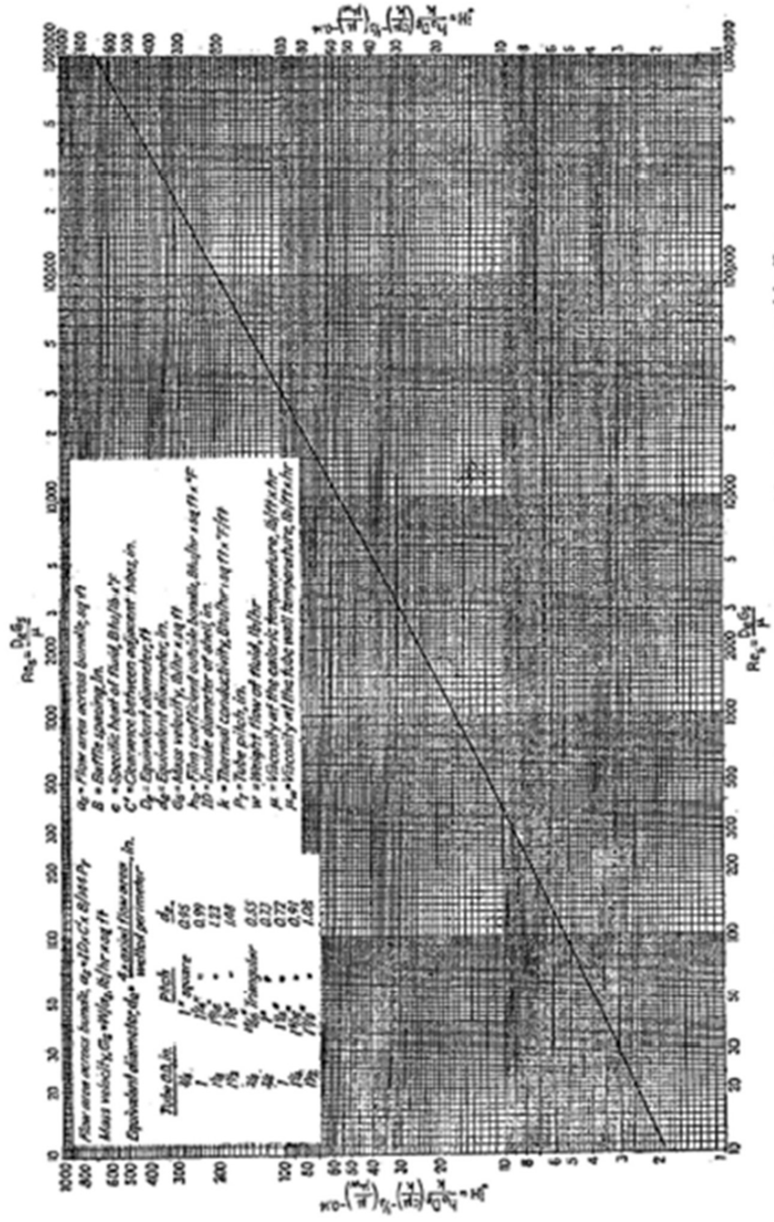

Fig. 28. Shell-side heat-transfer curve for bundles with 25% cut segmental baffle.

TABLE 10. HEAT EXCHANGER AND CONDENSER TUBE DATA

Tube OD, in.	BWG	Wall thickness, in.	ID, in.	Flow area per tube, in.²	Surface per lin ft, ft² Outside	Surface per lin ft, ft² Inside	Weight per lin ft, lb steel
½	12	0.109	0.282	0.0625	0.1309	0.0748	0.493
	14	0.083	0.334	0.0876		0.0874	0.403
	16	0.065	0.370	0.1076		0.0969	0.329
	18	0.049	0.402	0.127		0.1052	0.258
	20	0.035	0.430	0.145		0.1125	0.190
¾	10	0.134	0.482	0.182	0.1963	0.1263	0.965
	11	0.120	0.510	0.204		0.1335	0.884
	12	0.109	0.532	0.223		0.1393	0.817
	13	0.095	0.560	0.247		0.1466	0.727
	14	0.083	0.584	0.268		0.1529	0.647
	15	0.072	0.606	0.289		0.1587	0.571
	16	0.065	0.620	0.302		0.1623	0.520
	17	0.058	0.634	0.314		0.1660	0.469
	18	0.049	0.652	0.334		0.1707	0.401
1	8	0.165	0.670	0.355	0.2618	0.1754	1.61
	9	0.148	0.704	0.389		0.1843	1.47
	10	0.134	0.732	0.421		0.1916	1.36
	11	0.120	0.760	0.455		0.1990	1.23
	12	0.109	0.782	0.479		0.2048	1.14
	13	0.095	0.810	0.515		0.2121	1.00
	14	0.083	0.834	0.546		0.2183	0.890
	15	0.072	0.856	0.576		0.2241	0.781
	16	0.065	0.870	0.594		0.2277	0.710
	17	0.058	0.884	0.613		0.2314	0.639
	18	0.049	0.902	0.639		0.2361	0.545
1¼	8	0.165	0.920	0.665	0.3271	0.2409	2.09
	9	0.148	0.954	0.714		0.2498	1.91
	10	0.134	0.982	0.757		0.2572	1.75
	11	0.120	1.01	0.800		0.2644	1.58
	12	0.109	1.03	0.836		0.2701	1.45
	13	0.095	1.06	0.884		0.2775	1.28
	14	0.083	1.08	0.923		0.2839	1.13
	15	0.072	1.11	0.960		0.2896	0.991
	16	0.065	1.12	0.985		0.2932	0.900
	17	0.058	1.13	1.01		0.2969	0.808
	18	0.049	1.15	1.04		0.3015	0.688
1½	8	0.165	1.17	1.075	0.3925	0.3063	2.57
	9	0.148	1.20	1.14		0.3152	2.34
	10	0.134	1.23	1.19		0.3225	2.14
	11	0.120	1.26	1.25		0.3299	1.98
	12	0.109	1.28	1.29		0.3356	1.77
	13	0.095	1.31	1.35		0.3430	1.56
	14	0.083	1.33	1.40		0.3492	1.37
	15	0.072	1.36	1.44		0.3555	1.20
	16	0.065	1.37	1.47		0.3587	1.09
	17	0.058	1.38	1.50		0.3623	0.978
	18	0.049	1.40	1.54		0.3670	0.831