UNIVERSITY OF JOHANNESBURG

PROGRAM	BACCALAUREUS TECHNOLOGIAE *CHEMICAL ENGINEERING*
SUBJECT	**CHEMICAL ENGINEERING TECHNOLOGY IV (HEAT AND MASS)**
CODE	**WARC432**
DATE	MID YEAR EXAMINATION July 2019
DURATION	3HRS (X-PAPER) 08:30 – 11:30
TOTAL MARKS	119
FULL MARKS	119

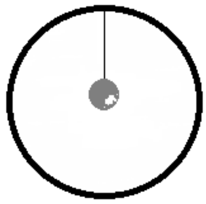
EXAMINERS	Prof M Belaid &Dr R Huberts	
MODERATOR	Prof R Mbaya	2366C
NUMBER OF PAGES	13	

INSTRUCTIONS	NON-PROGRAMMABLE CALCULATORS PERMITTED (ONLY ONE PER CANDIDATE) ANSWER ALL THE QUESTIONS.

QUESTION 1

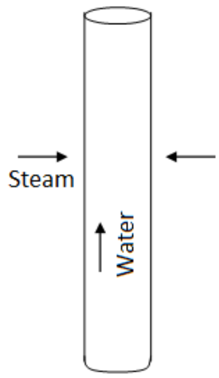
1.1. Show that $A_2F_{21} = A_1F_{12}$ for two infinitely large parallel flat surfaces of unequal size. (6)

1.2. A round black body ball of diameter 0.1m at an initial temperature of 527°C is suspended in the centre of a spherical oven of diameter 1m, filled with air at a temperature of 677°C and lined on the inside with brick at a temperature of 1027°C. The brick lining exhibits the same non-grey body characteristics as kaolin. Calculate the ***net*** radiation heat transfer ***to*** the ball. Show view factors in your calculations. (15)



[21]

QUESTION 2



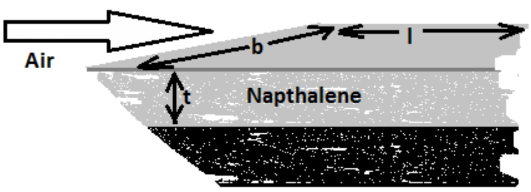
Steam at 1 atmosphere needs to condense at a rate of $0.001 kgs^{-1}$ on the outside of a tube with inside and outside diameter of 0.10m and 0.12m respectively and length 2m. The (turbulent) water flowing inside the tube is at 330K (average T_m) to provide the cooling, and the thermal resistance of the pipe is negligible.

2.1. What should the temperature of tube surface be? (35)

2.2. What should the Reynold's number (Re) of the cooling water inside the tube be to ensure such a surface temperature? (18)

[53]

QUESTION 3



A flat slab of naphthalene of dimensions 20cm × 10cm × 1cm (l×b×t) is exposed at the top to (fresh) air flowing at $12 ms^{-1}$. The MW naphthalene = $128.16 gmol^{-1}$, and the vapour pressure of naphthalene for the operating conditions (Total pressure = 1atm and 27°C) is 11.9Pa, with $D_{AB}=0.62\times10^{-5} m^2s^{-1}$. Mass transfer from the ends of the slab may be neglected.

3.1. Draw a diagram of the naphthalene concentration profile above the plate.(10)

3.2. Estimate the mass loss (g) of naphthalene after 24 hours of exposure to the air. (35)

[45]

TOTAL MARKS = 119

FULL MARKS = 119

DATA SHEETS

Lengths, areas and volumes:

Circumference of a circle πd

Area of a cylinder πdL

Area of a circle $\pi d^2/4$

Surface area of a sphere πd^2

Volume of a sphere $\pi d^3/6$

Absorbtivity and emissivity of Kaolin insulating brick:

800K ε = 0.70

1200K ε = 0.57

1400K ε = 0.47

1600K ε = 0.53

Thermal conductivity of Kaolin insulating brick:

$2 Wm^{-1}K^{-1}$

$\varepsilon_1 = \varepsilon_1(T_1)$ and $\alpha_1 = \varepsilon_1(T_2)$

TABLE A.4 Thermophysical Properties of Gases at Atmospheric Pressure[a]

T (K)	ρ (kg/m^3)	c_p (kJ/kg · K)	$\mu \cdot 10^7$ (N · s/m^2)	$\nu \cdot 10^6$ (m^2/s)	$k \cdot 10^3$ (W/m · K)	$\alpha \cdot 10^6$ (m^2/s)	Pr
Air							
100	3.5562	1.032	71.1	2.00	9.34	2.54	0.786
150	2.3364	1.012	103.4	4.426	13.8	5.84	0.758
200	1.7458	1.007	132.5	7.590	18.1	10.3	0.737
250	1.3947	1.006	159.6	11.44	22.3	15.9	0.720
300	1.1614	1.007	184.6	15.89	26.3	22.5	0.707
350	0.9950	1.009	208.2	20.92	30.0	29.9	0.700
400	0.8711	1.014	230.1	26.41	33.8	38.3	0.690
450	0.7740	1.021	250.7	32.39	37.3	47.2	0.686
500	0.6964	1.030	270.1	38.79	40.7	56.7	0.684
550	0.6329	1.040	288.4	45.57	43.9	66.7	0.683
600	0.5804	1.051	305.8	52.69	46.9	76.9	0.685
650	0.5356	1.063	322.5	60.21	49.7	87.3	0.690
700	0.4975	1.075	338.8	68.10	52.4	98.0	0.695
750	0.4643	1.087	354.6	76.37	54.9	109	0.702
800	0.4354	1.099	369.8	84.93	57.3	120	0.709
850	0.4097	1.110	384.3	93.80	59.6	131	0.716
900	0.3868	1.121	398.1	102.9	62.0	143	0.720
950	0.3666	1.131	411.3	112.2	64.3	155	0.723
1000	0.3482	1.141	424.4	121.9	66.7	168	0.726
1100	0.3166	1.159	449.0	141.8	71.5	195	0.728
1200	0.2902	1.175	473.0	162.9	76.3	224	0.728
1300	0.2679	1.189	496.0	185.1	82	238	0.719
1400	0.2488	1.207	530	213	91	303	0.703
1500	0.2322	1.230	557	240	100	350	0.685
1600	0.2177	1.248	584	268	106	390	0.688

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, $Pr \gtrsim 0.6$
$Nu_D = 3.66$	(8.55)	Laminar, fully developed, uniform T_s, $Pr \gtrsim 0.6$
$\overline{Nu}_D = 3.66 + \dfrac{0.0668(D/L)Re_D Pr}{1 + 0.04[(D/L)Re_D Pr]^{2/3}}$	(8.56)	Laminar, thermal entry length ($Pr \gg 1$ or an unheated starting length), uniform T_s
or $\overline{Nu}_D = 1.86\left(\dfrac{Re_D Pr}{L/D}\right)^{1/3}\left(\dfrac{\mu}{\mu_s}\right)^{0.14}$	(8.57)	Laminar, combined entry length $\{[Re_D Pr/(L/D)]^{1/3}(\mu/\mu_s)^{0.14}\} \gtrsim 2$, uniform T_s, $0.48 < Pr < 16{,}700$, $0.0044 < (\mu/\mu_s) < 9.75$
$f = 0.316 Re_D^{-1/4}$	(8.20a)[c]	Turbulent, fully developed, $Re_D \lesssim 2 \times 10^4$
$f = 0.184 Re_D^{-1/5}$	(8.20b)[c]	Turbulent, fully developed, $Re_D \gtrsim 2 \times 10^4$
or $f = (0.790 \ln Re_D - 1.64)^{-2}$	(8.21)[c]	Turbulent, fully developed, $3000 \lesssim Re_D \lesssim 5 \times 10^6$
$Nu_D = 0.023 Re_D^{4/5} Pr^n$	(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 $T_s < T_m$
or $Nu_D = 0.027 Re_D^{4/5} Pr^{1/3}\left(\dfrac{\mu}{\mu_s}\right)^{0.14}$	(8.61)[d]	Turbulent, fully developed, $0.7 \lesssim Pr \lesssim 16{,}700$, $Re_D \gtrsim 10{,}000$, $L/D \gtrsim 10$
or $Nu_D = \dfrac{(f/8)(Re_D - 1000)Pr}{1 + 12.7(f/8)^{1/2}(Pr^{2/3} - 1)}$	(8.63)[d]	Turbulent, fully developed, $0.5 < Pr < 2000$, $3000 \lesssim Re_D \lesssim 5 \times 10^6$, $(L/D) \gtrsim 10$
$Nu_D = 4.82 + 0.0185(Re_D Pr)^{0.827}$	(8.65)	Liquid metals, turbulent, fully developed, uniform q''_s, $3.6 \times 10^3 < Re_D < 9.05 \times 10^5$, $10^2 < Pe_D < 10^4$
$Nu_D = 5.0 + 0.025(Re_D Pr)^{0.8}$	(8.66)	Liquid metals, turbulent, fully developed, uniform T_s, $Pe_D > 100$

[a]The mass transfer correlations may be obtained by replacing Nu_D and Pr by Sh_D and Sc, respectively.

[b]Properties in Equations 8.53, 8.55, 8.60, 8.61, 8.63, 8.65, and 8.66 are based on $\overline{T}_m$; properties in Equations 8.19, 8.20, and 8.21 are based on $T_f \equiv (T_s + \overline{T}_m)/2$; properties in Equations 8.56 and 8.57 are based on $\overline{T}_m \equiv (T_{m,i} + T_{m,o})/2$.

[c]Equations 8.20 and 8.21 pertain to smooth tubes. For rough tubes, Equation 8.63 should be used with the results of Figure 8.3.

[d]As a first approximation, Equation 8.60, 8.61, or 8.63 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, $\overline{T}_m \equiv (T_{m,i} + T_{m,o})/2$.

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

Correlation		Geometry	Conditions
$\delta = 5x\,Re_x^{-1/2}$	(7.19)	Flat plate	Laminar, T_f
$C_{f,x} = 0.664Re_x^{-1/2}$	(7.20)	Flat plate	Laminar, local, T_f
$Nu_x = 0.332Re_x^{1/2}\,Pr^{1/3}$	(7.23)	Flat plate	Laminar, local, T_f, $0.6 \lesssim Pr \lesssim 50$
$\delta_t = \delta Pr^{-1/3}$	(7.24)	Flat plate	Laminar, T_f
$\overline{C}_{f,x} = 1.328Re_x^{-1/2}$	(7.30)	Flat plate	Laminar, average, T_f
$\overline{Nu}_x = 0.664Re_x^{1/2}\,Pr^{1/3}$	(7.31)	Flat plate	Laminar, average, T_f, $0.6 \lesssim Pr \lesssim 50$
$Nu_x = 0.565Pe_x^{1/2}$	(7.33)	Flat plate	Laminar, local, T_f, $Pr \lesssim 0.05$
$C_{f,x} = 0.0592Re_x^{-1/5}$	(7.35)	Flat plate	Turbulent, local, T_f, $Re_x \lesssim 10^8$
$\delta = 0.37x\,Re_x^{-1/5}$	(7.36)	Flat plate	Turbulent, local, T_f, $Re_x \lesssim 10^8$
$Nu_x = 0.0296Re_x^{4/5}\,Pr^{1/3}$	(7.37)	Flat plate	Turbulent, local, T_f, $Re_x \lesssim 10^8$, $0.6 \lesssim Pr \lesssim 60$
$\overline{C}_{f,L} = 0.074Re_L^{-1/5} - 1742Re_L^{-1}$	(7.43)	Flat plate	Mixed, average, T_f, $Re_{x,c} = 5 \times 10^5$, $Re_L \lesssim 10^8$
$\overline{Nu}_L = (0.037Re_L^{4/5} - 871)Pr^{1/3}$	(7.41)	Flat plate	Mixed, average, T_f, $Re_{x,c} = 5 \times 10^5$, $Re_L \lesssim 10^8$, $0.6 < Pr < 60$
$\overline{Nu}_D = C\,Re_D^m\,Pr^{1/3}$ (Table 7.2)	(7.55b)	Cylinder	Average, T_f, $0.4 < Re_D < 4 \times 10^5$, $Pr \gtrsim 0.7$
$\overline{Nu}_D = C\,Re_D^m\,Pr^n(Pr/Pr_s)^{1/4}$ (Table 7.4)	(7.56)	Cylinder	Average, T_∞, $1 < Re_D < 10^6$, $0.7 < Pr < 500$
$\overline{Nu}_D = 0.3 + [0.62Re_D^{1/2}\,Pr^{1/3}$ $\times [1 + (0.4/Pr)^{2/3}]^{-1/4}]$ $\times [1 + (Re_D/282{,}000)^{5/8}]^{4/5}$	(7.57)	Cylinder	Average, T_f, $Re_D\,Pr > 0.2$
$\overline{Nu}_D = 2 + (0.4Re_D^{1/2}$ $+ 0.06Re_D^{2/3})Pr^{0.4}$ $\times (\mu/\mu_s)^{1/4}$	(7.59)	Sphere	Average, T_∞, $3.5 < Re_D < 7.6 \times 10^4$, $0.71 < Pr < 380$
$\overline{Nu}_D = 2 + 0.6Re_D^{1/2}\,Pr^{1/3}$	(7.60)	Falling drop	Average, T_∞
$\overline{Nu}_D = 1.13C_1 Re_{D,\max}^m\,Pr^{1/3}$ (Tables 7.5, 7.6)	(7.63)	Tube bank[c]	Average, $\overline{T}_f$, $2000 < Re_{D,\max} < 4 \times 10^4$, $Pr \gtrsim 0.7$
$\overline{Nu}_D = C\,Re_{D,\max}^m\,Pr^{0.36}(Pr/Pr_s)^{1/4}$ (Tables 7.7, 7.8)	(7.67)	Tube bank[c]	Average, $\overline{T}$, $1000 < Re_D < 2 \times 10^6$, $0.7 < Pr < 500$
Single round nozzle	(7.79)	Impinging jet	Average, T_f, $2000 < Re < 4 \times 10^5$, $2 < (H/D) < 12$, $2.5 < (r/D) < 7.5$
Single slot nozzle	(7.82)	Impinging jet	Average, T_f, $3000 < Re < 9 \times 10^4$, $2 < (H/W) < 10$, $4 < (x/W) < 20$
Array of round nozzles	(7.84)	Impinging jet	Average, T_f, $2000 < Re < 10^5$, $2 < (H/D) < 12$, $0.004 < A_r < 0.04$
Array of slot nozzles	(7.87)	Impinging jet	Average, T_f, $1500 < Re < 4 \times 10^4$, $2 < (H/W) < 80$, $0.008 < A_r < 2.5A_{r,o}$
$\varepsilon\overline{j}_H = \varepsilon\overline{j}_m = 2.06Re_D^{-0.575}$	(7.91)	Packed bed of spheres[c]	Average, $\overline{T}$, $90 \lesssim Re_D \lesssim 4000$, $Pr \approx 0.7$

[a]Correlations in this table pertain to isothermal surfaces; for special cases involving an unheated starting length or a uniform surface heat flux, see Section 7.2.4.

[b]When the heat and mass transfer analogy is applicable, the corresponding mass transfer correlations may be obtained by replacing Nu and Pr by Sh and Sc, respectively.

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

TABLE 6.2 Selected dimensionless groups of heat and mass transfer

Group	Definition	Interpretation
Biot number (Bi)	$\dfrac{hL}{k_s}$	Ratio of the internal thermal resistance of a solid to the boundary layer thermal resistance.
Mass transfer Biot number (Bi_m)	$\dfrac{h_m L}{D_{AB}}$	Ratio of the internal species transfer resistance to the boundary layer species transfer resistance.
Bond number (Bo)	$\dfrac{g(\rho_l - \rho_v)L^2}{\sigma}$	Ratio of gravitational and surface tension forces.
Coefficient of friction (C_f)	$\dfrac{\tau_s}{\rho V^2/2}$	Dimensionless surface shear stress.
Eckert number (Ec)	$\dfrac{V^2}{c_p(T_s - T_\infty)}$	Kinetic energy of the flow relative to the boundary layer enthalpy difference.
Fourier number (Fo)	$\dfrac{\alpha t}{L^2}$	Ratio of the heat conduction rate to the rate of thermal energy storage in a solid. Dimensionless time.
Mass transfer Fourier number (Fo_m)	$\dfrac{D_{AB} t}{L^2}$	Ratio of the species diffusion rate to the rate of species storage. Dimensionless time.
Friction factor (f)	$\dfrac{\Delta p}{(L/D)(\rho u_m^2/2)}$	Dimensionless pressure drop for internal flow.
Grashof number (Gr_L)	$\dfrac{g\beta(T_s - T_\infty)L^3}{\nu^2}$	Ratio of buoyancy to viscous forces.
Colburn j factor (j_H)	$St\,Pr^{2/3}$	Dimensionless heat transfer coefficient.
Colburn j factor (j_m)	$St_m\,Sc^{2/3}$	Dimensionless mass transfer coefficient.
Jakob number (Ja)	$\dfrac{c_p(T_s - T_{sat})}{h_{fg}}$	Ratio of sensible to latent energy absorbed during liquid–vapor phase change.
Lewis number (Le)	$\dfrac{\alpha}{D_{AB}}$	Ratio of the thermal and mass diffusivities.
Nusselt number (Nu_L)	$\dfrac{hL}{k_f}$	Dimensionless temperature gradient at the surface.
Peclet number (Pe_L)	$\dfrac{VL}{\alpha} = Re_L Pr$	Dimensionless independent heat transfer parameter.
Prandtl number (Pr)	$\dfrac{c_p \mu}{k} = \dfrac{\nu}{\alpha}$	Ratio of the momentum and thermal diffusivities.
Reynolds number (Re_L)	$\dfrac{VL}{\nu}$	Ratio of the inertia and viscous forces.
Schmidt number (Sc)	$\dfrac{\nu}{D_{AB}}$	Ratio of the momentum and mass diffusivities.
Sherwood number (Sh_L)	$\dfrac{h_m L}{D_{AB}}$	Dimensionless concentration gradient at the surface.
Stanton number (St)	$\dfrac{h}{\rho V c_p} = \dfrac{Nu_L}{Re_L Pr}$	Modified Nusselt number.
Mass transfer Stanton number (St_m)	$\dfrac{h_m}{V} = \dfrac{Sh_L}{Re_L Sc}$	Modified Sherwood number.
Weber number (We)	$\dfrac{\rho V^2 L}{\sigma}$	Ratio of inertia to surface tension forces.

TABLE A.6 Thermophysical Properties of Saturated Water[a]

Temperature, T (K)	Pressure, P (bars)[b]	Specific Volume (m³/kg)		Heat of Vaporization, h_{fg} (kJ/kg)	Specific Heat (kJ/kg · K)		Viscosity (N · s/m²)		Thermal Conductivity (W/m · K)		Prandtl Number		Surface Tension, $\sigma_f \cdot 10^3$ (N/m)	Expansion Coefficient, $\beta_f \cdot 10^6$ (K⁻¹)	Temperature, T (K)
		$v_f \cdot 10^3$	v_g		$c_{p,f}$	$c_{p,g}$	$\mu_f \cdot 10^6$	$\mu_g \cdot 10^6$	$k_f \cdot 10^3$	$k_g \cdot 10^3$	Pr_f	Pr_g			
273.15	0.00611	1.000	206.3	2502	4.217	1.854	1750	8.02	569	18.2	12.99	0.815	75.5	−68.05	273.15
275	0.00697	1.000	181.7	2497	4.211	1.855	1652	8.09	574	18.3	12.22	0.817	75.3	−32.74	275
280	0.00990	1.000	130.4	2485	4.198	1.858	1422	8.29	582	18.6	10.26	0.825	74.8	46.04	280
285	0.01387	1.000	99.4	2473	4.189	1.861	1225	8.49	590	18.9	8.81	0.833	74.3	114.1	285
290	0.01917	1.001	69.7	2461	4.184	1.864	1080	8.69	598	19.3	7.56	0.841	73.7	174.0	290
295	0.02617	1.002	51.94	2449	4.181	1.868	959	8.89	606	19.5	6.62	0.849	72.7	227.5	295
300	0.03531	1.003	39.13	2438	4.179	1.872	855	9.09	613	19.6	5.83	0.857	71.7	276.1	300
305	0.04712	1.005	29.74	2426	4.178	1.877	769	9.29	620	20.1	5.20	0.865	70.9	320.6	305
310	0.06221	1.007	22.93	2414	4.178	1.882	695	9.49	628	20.4	4.62	0.873	70.0	361.9	310
315	0.08132	1.009	17.82	2402	4.179	1.888	631	9.69	634	20.7	4.16	0.883	69.2	400.4	315
320	0.1053	1.011	13.98	2390	4.180	1.895	577	9.89	640	21.0	3.77	0.894	68.3	436.7	320
325	0.1351	1.013	11.06	2378	4.182	1.903	528	10.09	645	21.3	3.42	0.901	67.5	471.2	325
330	0.1719	1.016	8.82	2366	4.184	1.911	489	10.29	650	21.7	3.15	0.908	66.6	504.0	330
335	0.2167	1.018	7.09	2354	4.186	1.920	453	10.49	656	22.0	2.88	0.916	65.8	535.5	335
340	0.2713	1.021	5.74	2342	4.188	1.930	420	10.69	660	22.3	2.66	0.925	64.9	566.0	340
345	0.3372	1.024	4.683	2329	4.191	1.941	389	10.89	668	22.6	2.45	0.933	64.1	595.4	345
350	0.4163	1.027	3.846	2317	4.195	1.954	365	11.09	668	23.0	2.29	0.942	63.2	624.2	350
355	0.5100	1.030	3.180	2304	4.199	1.968	343	11.29	671	23.3	2.14	0.951	62.3	652.3	355
360	0.6209	1.034	2.645	2291	4.203	1.983	324	11.49	674	23.7	2.02	0.960	61.4	697.9	360
365	0.7514	1.038	2.212	2278	4.209	1.999	306	11.69	677	24.1	1.91	0.969	60.5	707.1	365
370	0.9040	1.041	1.861	2265	4.214	2.017	289	11.89	679	24.5	1.80	0.978	59.5	728.7	370
373.15	1.0133	1.044	1.679	2257	4.217	2.029	279	12.02	680	24.8	1.76	0.984	58.9	750.1	373.15
375	1.0815	1.045	1.574	2252	4.220	2.036	274	12.09	681	24.9	1.70	0.987	58.6	761	375
380	1.2869	1.049	1.337	2239	4.226	2.057	260	12.29	683	25.4	1.61	0.999	57.6	788	380
385	1.5233	1.053	1.142	2225	4.232	2.080	248	12.49	685	25.8	1.53	1.004	56.6	814	385

Heat absorbed

$$Q = mC_p \Delta T$$

Fourier's law in one dimension:

$$q_x^{"} = -k \frac{dT}{dx}$$

Newton's law of cooling:

$q_x^{"} = h(T_S - T_\infty)$ use ΔT_{LM} for varying $(T_S - T_\infty)$

Stefan Boltzmann Law:

$$q^{"} = \sigma T_S^4$$

$$q^{"} = \varepsilon \sigma T_S^4$$

$$G = F_{Surr\ S} \alpha \sigma T_{Surr}^4$$

$$\sigma = 5.67 \times 10^{-8} Wm^{-2} K^4$$

Thermal diffusivity:

$$\alpha = \frac{k}{\rho C_p}$$

Average heat transfer coefficient for a generic surface and a flat plate:

$$\bar{h} = \frac{1}{A_s} \int_{A_s} h.dA_s$$

$$\bar{h} = \frac{1}{L} \int_0^L h.dx$$

Heat transfer at a boiling surface:

$q_S^{"} = h(T_S - T_{sat}) = h\Delta T_e$ …..10.3

Nucleate boiling correlation:

$q_S^{"} = \mu_l h_{fg} \left[\frac{g(\rho_l - \rho_v)}{\sigma} \right]^{1/2} \left(\frac{c_{p,l} \Delta T_e}{C_{s,f} h_{fg} \Pr_l^n} \right)^3$ **……10.5**

TABLE 1.1 Typical values of the convection heat transfer coefficient

Process	h ($W/m^2 \cdot K$)
Free convection	
Gases	2–25
Liquids	50–1000
Forced convection	
Gases	25–250
Liquids	100–20,000
Convection with phase change	
Boiling or condensation	2500–100,000

TABLE 10.1 Values of $C_{s,f}$ for various surface–fluid combinations [5–7]

Surface–Fluid Combination	$C_{s,f}$	n
Water–copper		
Scored	0.0068	1.0
Polished	0.0130	1.0
Water–stainless steel		
Chemically etched	0.0130	1.0
Mechanically polished	0.0130	1.0
Ground and polished	0.0060	1.0
Water–brass	0.0060	1.0
Water–nickel	0.006	1.0
Water–platinum	0.0130	1.0
n-Pentane–copper		
Polished	0.0154	1.7
Lapped	0.0049	1.7
Benzene–chromium	0.0101	1.7
Ethyl alcohol–chromium	0.0027	1.7

Critical heat flux (Kutateladze and Zuber):

$$q''_{\max} = 0.149 h_{fg} \rho_v \left[\frac{\sigma g (\rho_l - \rho_v)}{\rho_v^2} \right]^{1/4} \quad \textbf{......10.7}$$

Minimum heat flux (Zuber) (moderate pressures):

$$q''_{\min} = 0.09 \rho_v h_{fg} \left[\frac{g\sigma (\rho_l - \rho_v)}{(\rho_l + \rho_v)^2} \right]^{1/4} \quad \textbf{......10.8}$$

Film pool boiling for Ts<300ºC (radiation component low):

$$\overline{Nu}_D = \frac{\overline{h}_{conv}.D}{k_v} = C.\left[\frac{g\sigma(\rho_l - \rho_v) h'_{fg} D^3}{\nu_v k_v (T_s - T_{sat})} \right]^{1/4} \quad \ldots 10.9$$

C = 0.62 for horizontal cylinders and C = 0.67 for spheres.

$$h'_{fg} = h_{fg} + 0.80 c_{p,v} (T_S - T_{sat})$$

Film pool boiling for Ts>300ºC:

$$\overline{h}^{4/3} = \overline{h}_{conv}^{4/3} + \overline{h}_{rad}.\overline{h}^{1/3}$$

$$if \ \overline{h}_{rad} \prec \overline{h}_{conv};$$

$$\overline{h} = \overline{h}_{conv} + \frac{3}{4}\overline{h}_{rad} \qquad \textbf{.......10.10}$$

$$where \ \overline{h}_{rad} = \frac{\varepsilon\sigma (T_s^4 - T_{sat}^4)}{T_s - T_{sat}}$$

External forced convection boiling of a cylinder in cross flow:

High velocity: $\dfrac{q''_{\max}}{\rho_v h_{fg} V} = \dfrac{1}{\pi}\left[1 + \left(\dfrac{4}{We_D} \right)^{1/3} \right]$

Low velocity: $\dfrac{q''_{\max}}{\rho_v h_{fg} V} = \dfrac{\left(\rho_l / \rho_v \right)^{3/4}}{169\pi} + \dfrac{\left(\rho_l / \rho_v \right)^{1/2}}{19.2\pi We_D^{1/3}}$

$$We_D = \frac{\rho V^2 D}{\sigma}$$

If $\dfrac{q''_{\max}}{\rho_v h_{fg} V} \le \left[\dfrac{0.275}{\pi}\right]\left[\dfrac{\rho l}{\rho v}\right]^{1/2} + 1$ **then the velocity is high, otherwise it is low**

CONDENSATION:

Nusselt laminar film condensation correlation:

$$\overline{N}u_L = \frac{\overline{h}_L.L}{k_L} = 0.943\left[\frac{\rho_L g(\rho_L - \rho_V)h'_{fg}.L^3}{\mu_L.k_L(T_{sat} - T_s)}\right]^{1/4}$$ **…..10.31**

Modified latent heat of formation term for condensation (Rohsenow):

$h'_{fg} = h_{fg}(1 + 0.68Ja)$ **……10.26**

Total heat transfer to the surface:

$q = \bar{h}_L\, A(T_{sat} - T_S)$ **…..10.32**

Total condensation rate:

$\dot{m} = \dfrac{q}{h'_{fg}} = \dfrac{\overline{h}_L.A.(T_{sat} - T_s)}{h'_{fg}}$ **…..10.33**

Reynolds number for condensation

$\mathrm{Re}_\delta = \dfrac{4\dot{m}}{\mu_L.b} = \dfrac{4\rho_l u_m \delta}{\mu_L}$ **…..10.35**

Condensate mass flowrate:

$$\dot{m}(x) = \rho_l u_m b\delta(x)$$

$$\dot{m}(x) = b\frac{g\rho_l(\rho_l - \rho_v)\delta(x)^3}{3\mu_l}$$ …..10.19

u_m = mean velocity, δ = thickness, b = breadth of plate

Thickness of condensate:

$$\delta(x) = \left[\frac{4k_l\mu_l(T_{sat} - T_S)x}{g\rho_l(\rho_l - \rho_v)h_{fg}}\right]^{1/4}$$

$\mathrm{Re}_\delta \le 30$ for wave-free laminar flow:

$\dfrac{\overline{h}_L(\nu_l^2/g)^{1/3}}{k_l} = 1.47\,\mathrm{Re}_\delta^{-1/3}$ **…..10.37**

$30 \le \mathrm{Re}_\delta \le 1800$ for wavy laminar flow:

$$\nu_l = \frac{\mu_l}{\rho_l}$$

$$\frac{\overline{h}_L(\nu_l^2/g)^{1/3}}{k_l} = \frac{\mathrm{Re}_\delta}{1.08\,\mathrm{Re}_\delta^{1.22} - 5.2}$$

Re_δ > 1800 for turbulent flow:

$$\frac{\bar{h}_L (\nu_l^2 / g)^{1/3}}{k_l} = \frac{\mathrm{Re}_\delta}{8750 + 58\,\mathrm{Pr}^{-0.5}(\mathrm{Re}_\delta^{0.75} - 253)}$$

Radiation Data:

Stefan Boltzmann constant = 5.67 x 10^{-8}Wm^{-2}K^{-1}

View factors:

-Two parallel infinite surfaces:

$$F_{12} = F_{21} = 1$$

-Concentric cylinders, spheres and in general:

$F_{21}A_2 = F_{12}A_1$ **(Reciprocity Theorem)**

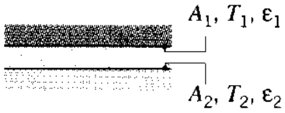
For metals: $\boldsymbol{\varepsilon_1 = \varepsilon_1(T_1)}$ **and** $\boldsymbol{\alpha_1 = \varepsilon_1(\sqrt{T_1 T_2})}$

TABLE 1.5 Summary of heat transfer processes

Mode	Mechanism(s)	Rate Equation	Equation Number	Transport Property or Coefficient
Conduction	Diffusion of energy due to random molecular motion	$q''_x\ (\mathrm{W/m^2}) = -k\dfrac{dT}{dx}$	(1.1)	k (W/m · K)
Convection	Diffusion of energy due to random molecular motion plus energy transfer due to bulk motion (advection)	$q''\ (\mathrm{W/m^2}) = h(T_s - T_\infty)$	(1.3a)	h (W/m^2 · K)
Radiation	Energy transfer by electromagnetic waves	$q''(\mathrm{W/m^2}) = \varepsilon\sigma(T_s^4 - T_{sur}^4)$ or $q\ (\mathrm{W}) = h_r A(T_s - T_{sur})$	(1.7) (1.8)	ε h_r (W/m^2 · K)

Table 13.3 Special Diffuse, Gray, Two-Surface Enclosures

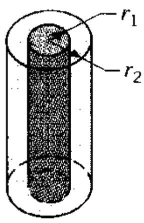
Large (Infinite) Parallel Planes



$$A_1 = A_2 = A \qquad F_{12} = 1$$

$$q_{12} = \frac{A\sigma(T_1^4 - T_2^4)}{\dfrac{1}{\varepsilon_1} + \dfrac{1}{\varepsilon_2} - 1} \tag{13.24}$$

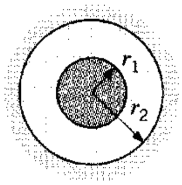
Long (Infinite) Concentric Cylinders



$$\frac{A_1}{A_2} = \frac{r_1}{r_2} \qquad F_{12} = 1$$

$$q_{12} = \frac{\sigma A_1(T_1^4 - T_2^4)}{\dfrac{1}{\varepsilon_1} + \dfrac{1 - \varepsilon_2}{\varepsilon_2}\left(\dfrac{r_1}{r_2}\right)} \tag{13.25}$$

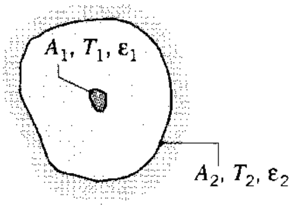
Concentric Spheres



$$\frac{A_1}{A_2} = \frac{r_1^2}{r_2^2} \qquad F_{12} = 1$$

$$q_{12} = \frac{\sigma A_1(T_1^4 - T_2^4)}{\dfrac{1}{\varepsilon_1} + \dfrac{1 - \varepsilon_2}{\varepsilon_2}\left(\dfrac{r_1}{r_2}\right)^2} \tag{13.26}$$

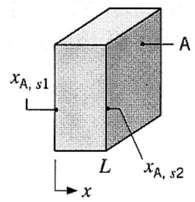
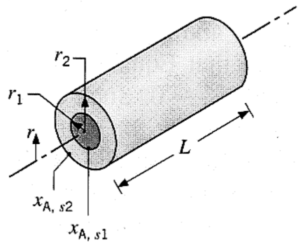
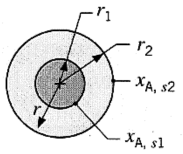
Small Convex Object in a Large Cavity



$$\frac{A_1}{A_2} \approx 0 \qquad F_{12} = 1$$

$$q_{12} = \sigma A_1 \varepsilon_1 (T_1^4 - T_2^4) \tag{13.27}$$

TABLE 14.1 Summary of Species Diffusion Solutions for Stationary Media with Specified Surface Concentrations[a]

Geometry	Species Concentration Distribution, $x_A(x)$ or $x_A(r)$	Species Diffusion Resistance, $R_{m,\,dif}$
	$x_A(x) = (x_{A,s2} - x_{A,s1})\dfrac{x}{L} + x_{A,s1}$	$R_{m,dif} = \dfrac{L}{D_{AB}A}$ [b]
	$x_A(r) = \dfrac{x_{A,s1} - x_{A,s2}}{\ln(r_1/r_2)}\ln\left(\dfrac{r}{r_2}\right) + x_{A,s2}$	$R_{m,dif} = \dfrac{\ln(r_2/r_1)}{2\pi L D_{AB}}$ [c]
	$x_A(r) = \dfrac{x_{A,s1} - x_{A,s2}}{1/r_1 - 1/r_2}\left(\dfrac{1}{r} - \dfrac{1}{r_2}\right) + x_{A,s2}$	$R_{m,\,dif} = \dfrac{1}{4\pi D_{AB}}\left(\dfrac{1}{r_1} - \dfrac{1}{r_2}\right)$ [c]

[a] Assuming C and D_{AB} are constant.
[b] $N_{A,x} = (C_{A,s1} - C_{A,s2})/R_{m,\,dif}$.
[c] $N_{A,r} = (C_{A,s1} - C_{A,s2})/R_{m,\,dif}$.

TABLE 3.3 One-dimensional, steady-state solutions to the heat equation with no generation

	Plane Wall	Cylindrical Wall[a]	Spherical Wall[a]
Heat equation	$\dfrac{d^2T}{dx^2} = 0$	$\dfrac{1}{r}\dfrac{d}{dr}\left(r\dfrac{dT}{dr}\right) = 0$	$\dfrac{1}{r^2}\dfrac{d}{dr}\left(r^2\dfrac{dT}{dr}\right) = 0$
Temperature distribution	$T_{s,1} - \Delta T\dfrac{x}{L}$	$T_{s,2} + \Delta T\dfrac{\ln(r/r_2)}{\ln(r_1/r_2)}$	$T_{s,1} - \Delta T\left[\dfrac{1-(r_1/r)}{1-(r_1/r_2)}\right]$
Heat flux (q'')	$k\dfrac{\Delta T}{L}$	$\dfrac{k\,\Delta T}{r\ln(r_2/r_1)}$	$\dfrac{k\,\Delta T}{r^2[(1/r_1) - (1/r_2)]}$
Heat rate (q)	$kA\dfrac{\Delta T}{L}$	$\dfrac{2\pi Lk\,\Delta T}{\ln(r_2/r_1)}$	$\dfrac{4\pi k\,\Delta T}{(1/r_1) - (1/r_2)}$
Thermal resistance ($R_{t,cond}$)	$\dfrac{L}{kA}$	$\dfrac{\ln(r_2/r_1)}{2\pi Lk}$	$\dfrac{(1/r_1) - (1/r_2)}{4\pi k}$

[a] The critical radius of insulation is $r_{cr} = k/h$ for the cylinder and $r_{cr} = 2k/h$ for the sphere.