Convective mass transfer

1 Errata

- Slide 10 gives 0.322 as the heat-column coefficient. Use 0.332.

  This agrees with the velocity and mass columns.

- Slide 12 gives 0.332 as the average laminar plate coefficient. Use 0.664 in the average Sherwood-number correlation

$$\overline{\mathrm{Sh}}_L = 0.664 \mathrm{Re}_L^{1/2} \mathrm{Sc}^{1/3}$$

- Slide 13 uses d/dx for the Reynolds-analogy gradients. Use d/dy.

  These gradients are normal to the wall.

- Slide 17 labels the subliming solid as carbon dioxide and supplies a vapour pressure of 4740 Pa without a surface temperature. Using the 293 K air temperature for solid CO_2 is physically inconsistent: NIST phase data place its triple point at 216.58 K. Its solid-vapour correlation gives 4740 Pa at about 163.5 K. Example 1 below instead uses a hypothetical volatile solid with explicitly stipulated properties and surface temperature. Its numerical answer must not be used for dry ice.

- Slide 20 omits the humidity and surface temperature needed for an evaporation rate. Example 4 below specifies dry air and a surface maintained at 35 degrees Celsius. An earlier worked solution used 6221 Pa at that temperature; use 5624 Pa. This pressure error was in the worked solution, rather than a printed value in the PDF question.

2 Reading list

- Çengel and Ghajar (2015) and Geankoplis (2003) for heat and mass transfer correlations.
- Bird et al. (2006) and Welty et al. (2001) for transport balances and diffusivity definitions.

Examples use mol unless stated otherwise.

3 Choose a convective-transfer model

Given conditions	Model and checks
Transfer between gas and liquid bulks	Use the two-film model, Eq. 1 and Eq. 2; put both resistances on the same basis.
A known interface concentration and one flowing phase	Use $N_A = k_c(c_{A,s} - c_{A,b})$ with the appropriate geometry and local or average Sherwood number.
A matching heat-transfer correlation is available	Use the heat/mass analogy only with matching geometry, flow and boundary conditions.
Bulk composition changes along a tube	Integrate a stream balance instead of using inlet concentration everywhere.
Buoyancy rather than imposed flow drives transfer	Choose a mass-buoyancy correlation and separate composition and temperature effects.

4 Deriving the two-film resistance

DEFINITION
Convective mass transfer

Species transport involving molecular diffusion and bulk fluid motion. The flow changes the concentration field and the gradient driving interfacial transfer.

Consider steady absorption of A from a gas into a liquid. A thin gas film and a thin liquid film supply the resistance. Each bulk phase is well mixed. Take $N_A > 0$ from gas to liquid, and neglect accumulation, reaction and interfacial resistance. The same molar flux crosses both films.

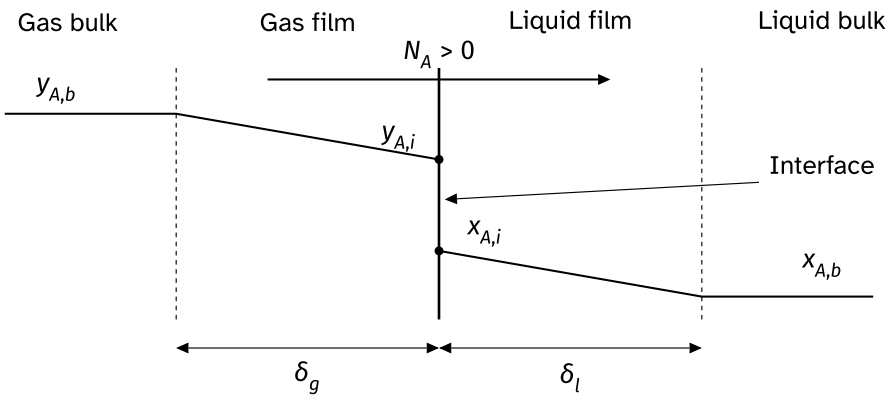


Figure 1: Two films meet at a gas-liquid interface. Positive species flux runs from the gas bulk through the gas film and liquid film into the liquid bulk.

The profile heights are schematic and use separate composition scales for the two phases. Equal chemical potential at the interface does not require equal gas and liquid mole fractions.

Let b mean bulk and i mean interface. Define the coefficients on a mole-fraction basis, in mol $\mathrm{m^{-2}\ s^{-1}}$ per unit mole-fraction difference:

$$\begin{aligned} N_A &= k_y(y_{A,b} - y_{A,i}) \\ N_A &= k_x(x_{A,i} - x_{A,b}) \\ y_{A,i} &= m x_{A,i}. \end{aligned} \tag{1}$$

The slope m comes from the local equilibrium relation. It is not a ratio of diffusivities. Solve the liquid-film equation for its interface value, then substitute into the gas-film equation:

$$\begin{aligned} x_{A,i} &= x_{A,b} + \frac{N_A}{k_x} \\ N_A &= k_y\left[y_{A,b} - m\left(x_{A,b} + \frac{N_A}{k_x}\right)\right] \\ y_{A,b} - m x_{A,b} &= N_A\left(\frac{1}{k_y} + \frac{m}{k_x}\right). \end{aligned}$$

Define the overall coefficient K_y using the gas-basis driving force:

$$\begin{aligned} N_A &= K_y(y_{A,b} - m x_{A,b}) \\ \frac{1}{K_y} &= \frac{1}{k_y} + \frac{m}{k_x} \end{aligned} \tag{2}$$

The factor m converts the liquid-film resistance to the gas basis. Comparing $1/k_y$ directly with $1/k_x$ would omit that conversion. A concentration-basis coefficient in m $\mathrm{s^{-1}}$ cannot be substituted without converting its driving force.

For example, take $k_y = 0.020$, $k_x = 0.010$ in the units above, $m = 2$, $y_{A,b} = 0.100$ and $x_{A,b} = 0.020$. Then

$$\begin{aligned} \frac{1}{K_y} &= 50 + 200 = 250\ \mathrm{m^2\,s\,mol^{-1}} \\ K_y &= 0.0040\ \mathrm{mol\,m^{-2}\,s^{-1}} \\ N_A &= 0.0040[0.100 - 2(0.020)] \\ &= 2.4 \times 10^{-4}\ \mathrm{mol\,m^{-2}\,s^{-1}}. \end{aligned}$$

The liquid film contributes 80 percent of the total gas-basis resistance. Recovering $x_{A,i} = 0.044$ and $y_{A,i} = 0.088$ gives the same flux through each film and satisfies $y_{A,i} = 2x_{A,i}$. Reducing m to 0.2 while keeping the coefficients fixed changes the resistances to 50 and 20; the gas film then controls most of the resistance. Solubility and the chosen basis matter.

QUIZ

1. Putting film resistances on one basis

Two-film transfer obeys $y_A^* = mx_A$. Which expression gives the overall gas-mole-fraction resistance?

1. $1/K_y = 1/k_y + 1/k_x$.
2. $1/K_y = 1/k_y + m/k_x$.
3. $K_y = k_y + k_x$.
4. $1/K_y = m/k_y + 1/k_x$.

See solution for Quiz 1

5 Deriving external-flow mass transfer

DEFINITION

Concentration boundary layer

The region near an interface where species concentration differs appreciably from its outer-fluid value. It has no sharp physical edge.

DEFINITION

Schmidt number

The dimensionless ratio $\mathrm{Sc} = \nu/D_{AB}$, where $\nu = \mu/\rho$ is the fluid's kinematic viscosity, or momentum diffusivity, and D_{AB} is the species diffusivity. Here μ is dynamic viscosity and ρ is fluid density.

For $\mathrm{Sc} \ll 1$, species diffuse faster than momentum and the concentration boundary layer is thicker than the velocity boundary layer. For $\mathrm{Sc} \gg 1$, species diffuse more slowly and the concentration layer is thinner. At $\mathrm{Sc} \approx 1$, the layers have comparable thicknesses under matching laminar-plate conditions. A liquid with $\nu = 10^{-6}\ \mathrm{m^2\,s^{-1}}$ and $D_{AB} = 10^{-9}\ \mathrm{m^2\,s^{-1}}$ has $\mathrm{Sc} = 1000$ and a thin concentration layer. Replacing species diffusivity by thermal diffusivity would give Prandtl number, which answers a different transport question.

DEFINITION

Lewis number

The dimensionless ratio

$$\mathrm{Le} = \frac{\alpha}{D_{AB}} = \frac{\mathrm{Sc}}{\mathrm{Pr}},$$

where α is the fluid's thermal diffusivity and D_{AB} is the species diffusivity, both in $\mathrm{m^2\,s^{-1}}$.

For $\mathrm{Le} \ll 1$, concentration disturbances diffuse faster than thermal disturbances. For $\mathrm{Le} \gg 1$, heat diffuses faster and the thermal layer extends farther than the concentration layer under matching flow conditions. At $\mathrm{Le} \approx 1$, matching heat and mass models can have similar scalar profiles, which supports a heat and mass transfer analogy when their boundary conditions also

match. For $\alpha = 10^{-7}\ \mathrm{m^2\,s^{-1}}$ and $D_{AB} = 10^{-9}\ \mathrm{m^2\,s^{-1}}$, $\mathrm{Le} = 100$; treating heat and species diffusion as equally fast would be a poor approximation. The air-water calculation below illustrates the near-unity case. Neither case permits setting $h = k_c$, since those coefficients have different units.

These diffusivity ratios help explain the relative thicknesses of the boundary layers. To calculate the mass transfer rate, we now need the concentration gradient at the surface. Consider fluid flowing over a wet flat plate, as shown in Figure 2. Let x measure distance along the plate in the flow direction, and let z measure perpendicular distance from the plate into the fluid, with $z = 0$ at the surface. At a fixed position x, the concentration falls from $c_{A,s}$ at the wet surface to $c_{A,\infty}$ in the free stream. The slope of this concentration profile at $z = 0$ gives the surface gradient in Fick's law. Below, we use that gradient to obtain the mass transfer coefficient.

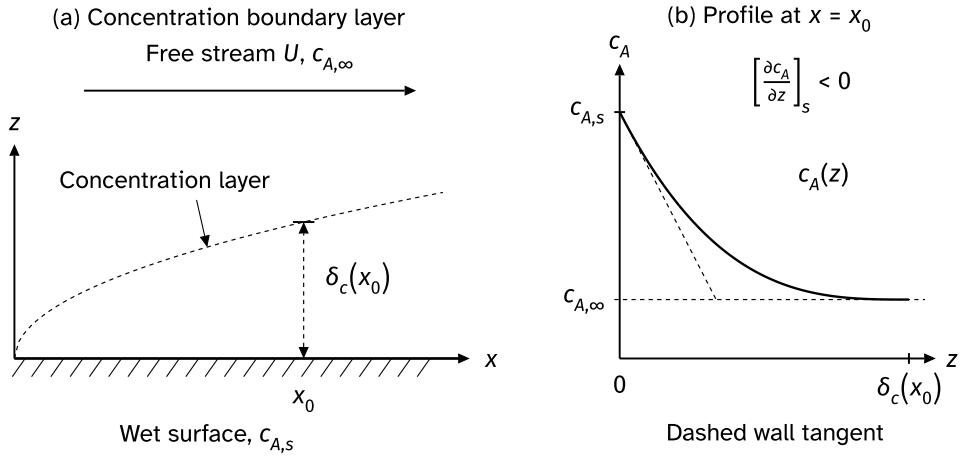


Figure 2: Concentration boundary layer over a wet plate and the local concentration profile at a marked downstream position. The layer edge is schematic; the dashed oblique tangent shows the negative wall gradient for evaporation.

5.1 The coefficient and Sherwood number

DEFINITION

Mass transfer coefficient

A coefficient relating species flux to a specified interfacial-to-bulk driving difference. On a molar-concentration basis,

$$N_A = k_c(c_{A,s} - c_{A,b})$$

and k_c has units $\mathrm{m\,s^{-1}}$; other driving-force bases use different units.

Use Sherwood number, first introduced in the spherical-diffusion calculation, to express the coefficient on a dimensionless basis. It corresponds to Nusselt number with species diffusivity replacing thermal conductivity in the appropriate coefficient definition.

Write the outgoing molar flux N_A in mol $\mathrm{m^{-2}\ s^{-1}}$ as a coefficient times a concentration difference. With constant total molar concentration c, its mole-fraction form is equivalent:

$$N_A = k_c(c_{A,s} - c_{A,\infty})$$

$$N_A = k_y(y_{A,s} - y_{A,\infty}), \quad k_y = ck_c \tag{3}$$

The concentration-basis coefficient k_c has units m s^{-1}. The source calls the mole-fraction coefficient k_x, but this topic uses k_y for gas composition and reserves k_x for liquid composition. Do not mix a coefficient defined per mole fraction with one defined per concentration.

At low normal mass flux, Fick's wall law is

$$N_A = -D_{AB}\left[\frac{\partial c_A}{\partial z}\right]_s.$$

Equating it with Eq. 3 gives

$$k_c = \frac{-D_{AB}\left[\frac{\partial c_A}{\partial z}\right]_s}{c_{A,s} - c_{A,\infty}} \tag{4}$$

The coefficient packages the wall gradient. Multiplication by a characteristic length L gives the dimensionless Sherwood number,

$$\mathrm{Sh} = \frac{k_c L}{D_{AB}}$$

$$\mathrm{Sh} = \frac{-\left[\frac{\partial c_A}{\partial z}\right]_s}{\frac{c_{A,s} - c_{A,\infty}}{L}} \tag{5}$$

It compares the actual gradient with a reference gradient over L. Large Sherwood means a steep wall gradient. It is analogous to $\mathrm{Nu} = hL/k$, where h is the heat-transfer coefficient in W m^{-2} K^{-1}. It is not a universal enhancement ratio with a stagnant value of one. A stagnant sphere has $k_c = D_{AB}/R$, hence $\mathrm{Sh} = 2$ when $L = 2R$.

QUIZ

2. From Sherwood number to coefficient

A correlation gives $\mathrm{Sh} = 100$ for $L = 0.10$ m and $D_{AB} = 2 \times 10^{-5}$ m^2 s^{-1}. What is k_c?

1. $0.20\ \mathrm{m\,s^{-1}}$.
2. $5\ \mathrm{m\,s^{-1}}$.
3. $2 \times 10^{-6}\ \mathrm{m\,s^{-1}}$.
4. $0.020\ \mathrm{m\,s^{-1}}$.

See solution for Quiz 2

5.2 Boundary-layer equations and local results

For steady two-dimensional laminar flow over a flat plate, assume constant properties, negligible streamwise diffusion, zero pressure gradient, no species reaction, no heat generation and low species injection. Continuity gives

$$\frac{\partial u}{\partial x}+\frac{\partial v}{\partial z}=0.$$

Here u and v are velocity components in m s^{-1}. The other balances are

$$\begin{aligned} uu_x+vu_z &= \nu u_{zz} \\ uT_x+vT_z &= \alpha T_{zz} \\ u(c_A)_x+v(c_A)_z &= D_{AB}(c_A)_{zz} \end{aligned} \tag{6}$$

Subscripts on variables in Eq. 6 denote derivatives. At $z=0$, impose $u=v=0$, $T=T_s$ and $c_A=c_{A,s}$. Far away, $u=U$, $T=T_\infty$ and $c_A=c_{A,\infty}$. The common advection operator and matching boundary conditions explain the analogy. If $\nu=\alpha=D_{AB}$, the normalized velocity, temperature and concentration profiles coincide. Blasius gives the wall velocity gradient

$$\frac{\partial u}{\partial z}\Big|_{z=0}=0.33206U\sqrt{\frac{U}{\nu x}}.$$

Thus, at $\mathrm{Sc}=1$, $\mathrm{Sh}_x=0.332\mathrm{Re}_x^{1/2}$ with $\mathrm{Re}_x=Ux/\nu$. The usual Schmidt correction gives

$$\begin{aligned} \mathrm{Sh}_x &= 0.332\mathrm{Re}_x^{1/2}\mathrm{Sc}^{1/3} \\ k_{c,x} &= 0.332D_{AB}\sqrt{\frac{U}{\nu x}}\mathrm{Sc}^{1/3} \end{aligned} \tag{7}$$

For a constant surface concentration the coefficient varies as $x^{-1/2}$. Its length average is

$$\bar{k}_c=L^{-1}\int_0^L k_{c,x}\,dx.$$

The integral is $2\sqrt{L}$, so the average is twice the trailing-edge value:

$$\overline{\mathrm{Sh}}_L=\frac{\bar{k}_cL}{D_{AB}}=0.664\mathrm{Re}_L^{1/2}\mathrm{Sc}^{1/3} \tag{8}$$

Use the laminar expression for $\mathrm{Re}_L<5\times10^5$ and $\mathrm{Sc}>0.5$.

5.3 Reynolds analogy

DEFINITION

Mass Stanton number

The dimensionless species-transfer coefficient

$$\mathrm{St}_D=\frac{k_c}{U}=\frac{\mathrm{Sh}}{\mathrm{Re}\,\mathrm{Sc}},$$

using a concentration-basis k_c in m s^{-1} and the same speed and length throughout. It compares surface transfer velocity with stream velocity and corresponds to the thermal Stanton number.

Small St_D means weak transfer relative to the flowing species capacity per area; larger values mean stronger transfer on that basis. For $k_c = 0.01\ \mathrm{m\,s^{-1}}$ and $U = 2\ \mathrm{m\,s^{-1}}$, $\mathrm{St}_D = 0.005$. This is not directly the fraction absorbed by a whole pipe, which also depends on length and diameter through Number of transfer units. A mole-fraction coefficient such as k_y cannot be divided by U here without converting its basis.

At $\mathrm{Pr} = 1$, normalize the profiles as

$$\frac{u}{U} = \frac{T - T_s}{T_\infty - T_s}.$$

Differentiate normal to the wall, multiply by $C_p\mu = k$, and recognize the wall shear $\tau_w = \mu u_z|_s$ and Fourier wall heat flux. This gives $h = C_p\tau_w/U$. Defining the skin-friction coefficient

$$C_f = \frac{\tau_w}{\frac{1}{2}\rho U^2}$$

gives

$$\frac{C_f}{2} = \mathrm{St}_H = \frac{h}{\rho U C_p}$$

$$\frac{C_f}{2} = \mathrm{St}_D = \frac{k_c}{U} \tag{9}$$

The second equality requires $\mathrm{Sc} = 1$ and matching transport conditions. The wall-normal gradients determine surface shear and heat flux. The dimensionless forms are

$$\mathrm{St}_H = \frac{\mathrm{Nu}}{\mathrm{RePr}}$$

and

$$\mathrm{St}_D = \frac{\mathrm{Sh}}{\mathrm{ReSc}}.$$

For a pipe the Fanning friction factor is the same shear normalization, while the Darcy factor is four times it. This distinction prevents a factor-of-four error. Do not replace wall skin friction by the total drag coefficient of a cylinder or sphere, where pressure drag can dominate. The simple analogy extends approximately to appropriate turbulent flows when momentum and scalar transport have comparable effective diffusivities; unity molecular ratios alone do not prove every turbulent profile identical.

5.4 Chilton-Colburn analogy and Lewis relation

DEFINITION
Colburn mass factor

The dimensionless factor $j_D = \mathrm{St}_D \mathrm{Sc}^{2/3}$, using concentration-basis mass Stanton number. In matching flows it can be compared with the Colburn heat factor, $j_H = \mathrm{St}_H \mathrm{Pr}^{2/3}$. Equality of the factors is an empirical analogy with geometry and transport restrictions.

The empirical transfer factors relax the unity-property condition:

$$j_H = \mathrm{St}_H \mathrm{Pr}^{2/3}$$

$$j_D = \mathrm{St}_D \mathrm{Sc}^{2/3}, \quad j_H \approx j_D \tag{10}$$

For suitable plates and pipes these factors also approximate $C_f/2$. The transfer-to-transfer relation can remain useful when total drag cannot be used, provided geometry, boundary conditions and transport assumptions match. It is not a guarantee for rough packed beds or high blowing. The source quotes approximate ranges $0.6 < \mathrm{Pr} < 60$ and $0.6 < \mathrm{Sc} < 3000$; individual correlations may impose tighter ranges. Equating factors and cancelling U yields

$$\frac{h}{k_c} = \rho C_p \left(\frac{\mathrm{Sc}}{\mathrm{Pr}} \right)^{2/3}$$

$$h = \rho C_p k_c \mathrm{Le}^{2/3} \tag{11}$$

With water vapour in air at 298 K, $D_{AB} = 2.5 \times 10^{-5}$ and $\alpha = 2.18 \times 10^{-5}$ m^2 s^{-1} give $\mathrm{Le} = 0.872$ and $\mathrm{Le}^{2/3} = 0.9127$. Dropping this roughly nine-percent correction gives the Lewis relation $h \approx \rho C_p k_c$. It supports wet-bulb and humidification models for air-water; other pairs need their own Lewis number.

QUIZ
3. The Lewis relation

The Chilton-Colburn heat and mass analogy applies, and $\mathrm{Le} = 1$. Which relation follows for concentration-basis k_c?

1. $h = \rho C_p k_c$.
2. $h = k_c$.
3. $h = \rho C_p / k_c$.
4. $h = k_c D_{AB}$.

See solution for Quiz 3

6 Classroom examples 1 to 5

All examples use mol rather than kmol. Multiply a concentration or rate in kmol by 1000 to obtain mol. The molar gas constant is $R = 8.314$ J mol^{-1} K^{-1}. The molar rate is $\dot{n}_A$ in mol s^{-1}

and mass rate is $\dot{m}_A = M_A \dot{n}_A$ in kg s^{-1}. An average coefficient carries a bar; A_s is transfer area in m^2.

EXERCISE

Exercise 1. Volatile solid

Air flows at 2 m s^{-1} over a hypothetical volatile solid A with an exposed area of $10^{-3}\ \mathrm{m}^2$. The measured sublimation rate is $2.29 \times 10^{-4}\ \mathrm{mol\,s^{-1}}$. Both the air and the surface are maintained at 293 K; external heating supplies the latent heat. The pressure is $1.013 \times 10^5\ \mathrm{Pa}$ and incoming air contains negligible A.

For this model solid, use the stipulated equilibrium vapour pressure $p_{A,s} = 4740\ \mathrm{Pa}$ and diffusivity in air $D_{AB} = 1.5 \times 10^{-5}\ \mathrm{m^2\,s^{-1}}$. Use air properties $\nu = 1.55 \times 10^{-5}\ \mathrm{m^2\,s^{-1}}$, $\mathrm{Pr} = 0.71$, $C_p = 1007\ \mathrm{J\,kg^{-1}\,K^{-1}}$ and molar mass $M_B = 0.029\ \mathrm{kg\,mol^{-1}}$.

Determine the concentration-basis mass transfer coefficient and estimate the heat transfer coefficient using the Chilton-Colburn analogy. This adaptation replaces the inconsistent CO_2 interface state in slide 17.

See solution for Exercise 1

EXERCISE
Exercise 2. Naphthalene sheet

A stream of air at 100 kPa pressure and 290 K is flowing on the top surface of a thin flat sheet of solid naphthalene of length 1 m with a velocity of 20 m/s and surface temperature of 310K. Calculate:

- The average mass transfer coefficient over the flat plate
- The rate of loss of naphthalene from the surface per unit width

Data @ $T_f = 300$ K:

- $D_{AB} = 6 \times 10^{-6}$ m^2/s (naphthalene vapour in air)
- $\nu = 1.5 \times 10^{-5}$ m^2/s (kinematic viscosity of air)
- $C_{As} = 1 \times 10^{-5}$ kmol/m^3 (conc. of naphthalene at the air-solid interface)

See solution for Exercise 2

EXERCISE
Exercise 3. Wet-bulb thermometer

Dry air at 1.013×10^5 Pa blows across a thermometer whose bulb has been covered with a dampened wick. The classical "wet bulb" thermometer indicates a steady-state temperature is reached by a small amount of liquid water evaporating into a large reservoir of unsaturated vapour-gas mixture. The thermometer reads 290 K. At this temperature, the following properties were evaluated:

Property	Value
Vapour pressure of water	1.94×10^3 Pa
Density of air	1.219 kg/m^3
Latent heat of vaporisation of water	2461 kJ/kg
Prandtl number	0.71
Schmidt number	0.61
Specific heat of air	1.006 kJ/kg.K

What is the temperature of the dry air?

See solution for Exercise 3

EXERCISE
Exercise 4. Heat and mass analogy

Water evaporates from a wetted surface of rectangular shape, with an air stream (1 atm, 35 °C) blown over the surface at a velocity of 15 m/s. Heat transfer measurements

indicate that for air at 1 atm and 35 °C, the average heat transfer coefficient is given by the following empirical relation:

$$\langle h \rangle = 21 V_\infty^{0.6}$$

where $\langle h \rangle$ is in W/(m^2 K) and V_∞ is in m/s. Estimate the mass transfer coefficient and evaporation rate for an area of 1.5 m^2. Assume dry incoming air and a surface maintained at 35 °C by external heating. Use $p_A^{\text{sat}} = 5624\,\text{Pa}$ at $T_s = 308.15\,\text{K}$ and the following rounded, stipulated transport properties: $\rho = 1.128\,\text{kg/m}^3$, $C_p = 1007\,\text{J/(kg K)}$, $\nu = 15.89 \times 10^{-6}\,\text{m}^2/\text{s}$, $D_{AB} = 2.6 \times 10^{-5}\,\text{m}^2/\text{s}$ and $\text{Pr} = 0.71$.

See solution for Exercise 4

EXERCISE

Exercise 5. Annular transfer

The following correlation has been proposed to characterise the heat transfer of fluids flowing in the annular section between concentric tubes:

$$\text{Nu}_{d_1} = \frac{h d_1}{k} = C \text{Re}_{d_1}^{0.8} \text{Pr}^{1/3} \left(\frac{\mu}{\mu_s} \right)^{0.14} \left(\frac{d_2}{d_1} \right)^{0.15} \tag{16}$$

Show how this equation can be used to predict the mass-transfer coefficient from a naphthalene rod that is used to form the inner tube of an annular duct. d_1 and d_2 are the diameters of the inner and outer tubes, respectively, whereas μ and μ_s are the viscosity of the fluid at the bulk and surface temperatures.

See solution for Exercise 5

7 Example: concentration development in a pipe

At a transferring pipe wall the concentration layer grows inward until it reaches the centre. Fully developed concentration means that the normalized profile no longer changes, even while the bulk concentration continues changing along the pipe. Hydrodynamic, thermal and concentration development have different lengths. In laminar flow the estimates are $L_h \approx 0.05 \text{Re}\, d$, $L_t \approx \text{Pr} L_h$ and $L_c \approx \text{Sc} L_h$, with d the pipe diameter. At $\text{Re} = 1000$, $\text{Sc} = 600$ and $d = 0.01$ m, $L_c \approx 300$ m. A short pipe then needs an entrance-region correlation. In turbulent flow, development is often of order ten diameters; check it rather than inferring development solely from Reynolds number.

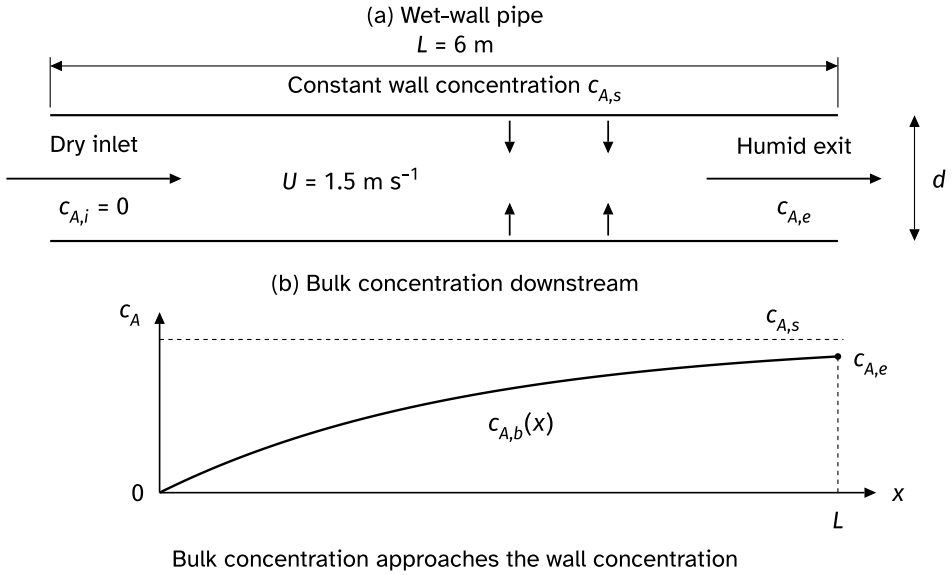


Figure 4: Water enters through a wet pipe wall at constant wall concentration. The separate schematic graph shows bulk concentration rising from its dry-inlet value toward the wall concentration as distance downstream increases.

QUIZ

4. Concentration development in a pipe

Water containing a slowly diffusing solute flows laminarly through a pipe with $\mathrm{Sc} = 600$. Using the entry-length estimates above, how does the concentration entry length compare with the hydrodynamic entry length?

1. It is approximately 600 times shorter.
2. It is approximately the same length.
3. It is approximately 600 times longer.
4. It is approximately 24 times longer.

See solution for Quiz 4

7.1 Translating a heat correlation

Use $\alpha \to D_{AB}$, $\mathrm{Pr} \to \mathrm{Sc}$ and $\mathrm{Nu} \to \mathrm{Sh}$ with the same geometry, wall condition and flow regime. The transfer must not appreciably modify the velocity field. The carrier fluid's properties approximate mixture properties only at dilution.

QUIZ

5. Matching a heat correlation

A heat correlation assumes uniform wall temperature and negligible normal injection. Which mass-transfer situation most closely matches its assumptions?

1. Only parts of the surface are wet, and strong flow out of the surface is ignored.
2. The wall concentration is uniform, flow through the wall is negligible, and the geometry and flow match.
3. Any reacting porous bed with unknown wetted area.
4. Any surface, provided Nusselt is renamed Sherwood.

See solution for Quiz 5

7.2 Natural convection and buoyancy

DEFINITION

Mass Grashof number

The composition-buoyancy group

$$\mathrm{Gr}_D = \frac{g|\Delta\rho_D|L^3}{\rho\nu^2},$$

where the density difference is due only to composition at a common temperature and pressure. Use the correlation's reference density and characteristic length.

Small Gr_D means weak composition buoyancy on the viscous scale; large values mean stronger natural motion is possible. The water-pan example below calculates Gr_D from density and uses $L = A_s/P_s$. Using pan diameter instead makes its value 64 times too large. As with thermal Grashof number, the cube of length makes that choice especially important. The sign and orientation of the density change decide whether buoyancy assists or opposes flow; a magnitude alone does not.

DEFINITION

Mass Rayleigh number

The composition natural-convection group

$$\mathrm{Ra}_D = \mathrm{Gr}_D\mathrm{Sc} = \frac{g|\Delta\rho_D|L^3}{\rho\nu D_{AB}}.$$

It combines composition buoyancy, momentum diffusion and species diffusion, using the same reference state and length in both factors.

Small Ra_D supports a diffusion-dominated picture on the model scale; larger values can support stronger convective transfer when the density arrangement drives motion. Check the geometry and stability before using a correlation. The water-pan example gives $\mathrm{Ra}_D = 9.96 \times 10^4$, within its stated upward-facing surface range. Multiplying Gr_D by Pr instead of Sc would give the wrong diffusion comparison. Thermal Rayleigh number uses Pr.

Density differences generate natural flow. Water vapour is lighter than air at the same temperature and pressure, so composition alone can make air above water buoyant. Thermal expansion can reinforce or oppose that change. The thermal and composition Grashof magnitudes are

$$\mathrm{Gr}_T = \frac{g\beta|T_s - T_\infty|L^3}{\nu^2}$$

$$\mathrm{Gr}_D = \frac{g|\Delta\rho_D|L^3}{\rho\nu^2} \tag{18}$$

Here g is gravity in m s^{-2}, β the expansion coefficient in K^{-1}, and $\Delta\rho_D$ the density difference caused only by composition, evaluated at a common temperature. Both groups are dimensionless and use L^3. A total density difference between states at different temperatures already includes thermal buoyancy and must not have a thermal group added again. For small density variations one can add signed buoyancy contributions referenced to the same state; opposing forces require a suitable coupled correlation. Heat and concentration profiles also have different diffusivities, so adding groups alone does not establish an exact combined Sherwood law. If a thermal natural-convection correlation supplies h, Eq. 11 gives a conditional mass estimate. If composition drives flow, use an appropriate Sherwood correlation directly.

QUIZ

6. Avoiding double-counted buoyancy

A measured density difference already includes both temperature and composition changes between two states. How should it be used in buoyancy modelling?

1. Add a thermal Grashof contribution to it automatically.
2. Discard composition effects whenever temperature differs.
3. Replace the density difference by the species diffusivity.
4. Use the measured total density difference once, including both temperature and composition effects.

See solution for Quiz 6

7.3 Correlations used in classroom examples 6 and 7

For developed turbulent pipe flow with fixed wall concentration, the slide correlation is

$$\mathrm{Sh} = 0.023\mathrm{Re}^{0.8}\mathrm{Sc}^{0.4},$$

with diameter in Re and Sh, $\mathrm{Re} > 10^4$ and $0.7 < \mathrm{Sc} < 160$.

For the upward-facing pan with light fluid above the surface,

$$\mathrm{Sh} = 0.54(\mathrm{Gr}_D\mathrm{Sc})^{1/4}$$

over $10^4 < \mathrm{Gr}_D\mathrm{Sc} < 10^7$. Use $L = A_s/P_s$, where P_s is the perimeter. The worked examples check these ranges.

8 Classroom example: internal pipe flow

EXERCISE
Exercise 6. Wet felt pipe

Dry air (310 K, 1.013×10^5 Pa, 1.5 m/s) is flowing through a pipe lined with a felt material that is continuously saturated with water at 290 K. The pipe is 0.15 m in diameter and 6 m long. The saturated felt material has a roughness index, e/D, of 0.00001. Assuming that the temperature of the air and pipe wall is constant, determine the rate at which water must be added to keep the felt continuously saturated.

See solution for Exercise 6

9 Classroom example: natural convection

EXERCISE
Exercise 7. Warm water pan

A 225 mm diameter pan of water is maintained at 37 °C is exposed to a dry, still atmosphere at 17 °C. Estimate the heat and mass transfer coefficients, as well as the evaporation rate from the pan,

a. Assuming the convection is driven by species transfer; b. Assuming that the convection is driven by temperature gradients.

See solution for Exercise 7

10 Additional examples

EXERCISE
Exercise 8. Evaporation from a tray and an outlet-humidity check

A water tray is flush with the floor of a rectangular duct. The tray is 0.60 m long along the air flow and 0.25 m wide. The duct is 0.25 m wide and 0.12 m high. Initially dry air enters at 2.0 m s^{-1}, 30 °C and 101325 Pa. A heater maintains the water at 30 °C. Assume a uniform air stream over the tray and a concentration boundary layer beginning at its leading edge. Neglect upstream velocity-profile effects.

Use $\nu = 1.60 \times 10^{-5}$ m^2 s^{-1}, $D_{AB} = 2.60 \times 10^{-5}$ m^2 s^{-1}, water saturation pressure 4240 Pa, $M_A = 0.0180$ kg mol^{-1} and $R = 8.314$ J mol^{-1} K^{-1}. Calculate the average mass transfer coefficient and evaporation rate. Estimate the outlet relative humidity and use it to check the assumption of dry bulk air over the entire tray. Use the dilute-transfer flat-plate model without a Stefan-flow correction.

See solution for Exercise 8

EXERCISE

Exercise 9. Drying a thin coating on a spherical pellet

A rigid spherical pellet of diameter 0.010 m has a thin wet coating containing 0.0020 g of volatile solvent per cm² of pellet surface. Solvent-free air flows past it at 0.10 m s⁻¹. Both the air and pellet surface are maintained at 300 K and 101325 Pa. External temperature control supplies the evaporation heat.

Use air properties $\rho = 1.18$ kg m⁻³, $\mu = 1.85 \times 10^{-5}$ Pa s, $k = 0.0263$ W m⁻¹ K⁻¹ and $\mathrm{Pr} = 0.707$. The solvent has $D_{AB} = 9.62 \times 10^{-6}$ m² s⁻¹, saturation pressure 1270 Pa at 300 K, molar mass 0.0780 kg mol⁻¹ and liquid density 876.5 kg m⁻³. For this low-Re sphere, adopt the matching correlations

$$\mathrm{Nu}_D = 2 + 0.6 \mathrm{Re}_D^{1/2} \mathrm{Pr}^{1/3}$$

$$\mathrm{Sh}_D = 2 + 0.6 \mathrm{Re}_D^{1/2} \mathrm{Sc}^{1/3}$$

Estimate the heat and mass transfer coefficients and the drying time. Assume the exposed area stays saturated at the stated vapour pressure until the solvent is exhausted, coating resistance is negligible, and normal evaporation flow is small. Check the thin-coating assumption.

See solution for Exercise 9

11 What you should now be able to calculate

Recover a mass-transfer coefficient, combine phase resistances, and calculate evaporation, drying or stream enrichment with the stated interfacial and bulk conditions.

Practice: Workshop 11 for coating removal and tube dissolution.

12 Report an error

Email Ranjeet Utikar with the title Convective mass transfer, the URL `/mass-transport/convective-mass-transfer/`, and the equation or passage to check.

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Solutions

Quiz solutions

Quiz 1. Putting film resistances on one basis

Correct choice: 2.

Use the equilibrium slope to express both film resistances in terms of gas mole fraction:

$$\frac{1}{K_y} = \frac{1}{k_y} + \frac{m}{k_x}$$

The factor multiplying the liquid-film resistance converts a liquid mole-fraction difference to the equivalent gas mole-fraction difference. Both resistances must use the same basis before they are added.

Back to Quiz 1

Quiz 2. From Sherwood number to coefficient

Correct choice: 4.

Rearrange the Sherwood-number definition:

$$k_c = \frac{\mathrm{Sh} D_{AB}}{L}$$

Substituting the given values,

$$k_c = \frac{100(2 \times 10^{-5})}{0.10} = 0.020\ \mathrm{m\,s^{-1}}$$

Use the species diffusivity in this mass transfer calculation.

Back to Quiz 2

Quiz 3. The Lewis relation

Correct choice: 1.

The heat and mass transfer analogy gives

$$h = \rho C_p k_c \mathrm{Le}^{2/3}$$

With Lewis number equal to one,

$$h = \rho C_p k_c$$

The density and heat capacity are still needed. The heat and mass transfer coefficients have different units and cannot be set equal.

Back to Quiz 3

Quiz 4. Concentration development in a pipe

Correct choice: 3.

The laminar entry-length estimates give

$$\frac{L_c}{L_h} \approx \mathrm{Sc} = 600$$

The concentration profile needs about 600 times as much distance to develop as the velocity profile. A developed velocity profile therefore does not guarantee a developed concentration profile.

Back to Quiz 4

Quiz 5. Matching a heat correlation

Correct choice: 2.

The analogy requires matching geometry, flow and boundary conditions as well as corresponding transport properties. Uniform temperature maps to uniform concentration. Strong injection changes the velocity field and needs a model that includes it.

Back to Quiz 5

Quiz 6. Avoiding double-counted buoyancy

Correct choice: 4.

A total density difference already includes thermal buoyancy. Adding a separate thermal contribution would count it again. If separating mechanisms, reference them to the same state and retain their signs because they can reinforce or oppose each other.

Back to Quiz 6

Exercise solutions

Exercise 1. Volatile solid

Data

Quantity	Value
Exposed area	$A_s = 10^{-3}\ \mathrm{m}^2$
Measured molar rate	$\dot{n}_A = 2.29 \times 10^{-4}\ \mathrm{mol\,s^{-1}}$
Gas temperature	$T = 293\ \mathrm{K}$
Gas pressure	$P = 101300\ \mathrm{Pa}$
Species diffusivity	$D_{AB} = 1.5 \times 10^{-5}\ \mathrm{m^2\,s^{-1}}$
Kinematic viscosity	$\nu = 1.55 \times 10^{-5}\ \mathrm{m^2\,s^{-1}}$
Surface equilibrium pressure	$p_{A,s} = 4740\ \mathrm{Pa}$
Prandtl number	$\mathrm{Pr} = 0.71$
Specific heat capacity	$C_p = 1007\ \mathrm{J\,kg^{-1}\,K^{-1}}$
Air molar mass	$M_B = 0.029\ \mathrm{kg\,mol^{-1}}$

These values are stipulated for the hypothetical solid and air.

Assumptions

- Steady sublimation over the exposed area.
- Negligible A in the bulk air.
- Ideal gas behaviour and constant supplied properties.
- Low normal mass flux and matching heat and mass transfer conditions.

Calculations

The gas-side surface concentration follows from ideal-gas equilibrium:

$$c_{A,s} = \frac{p_{A,s}}{RT} = \frac{4740}{8.314(293)} = 1.9458\ \mathrm{mol\,m^{-3}}$$

The measured rate equals coefficient times area times concentration driving force:

$$\dot{n}_A = k_c A_s (c_{A,s} - c_{A,\infty})$$

Since the bulk concentration is negligible, solve for k_c.

From Eq. 3,

$$k_c = \frac{2.29 \times 10^{-4}}{10^{-3} \times 1.9458} = 0.1177\ \mathrm{m\,s^{-1}}$$

The Schmidt number uses momentum and molecular diffusivities:

$$\mathrm{Sc} = \frac{\nu}{D_{AB}} = \frac{1.55 \times 10^{-5}}{1.5 \times 10^{-5}} = 1.0333$$

$$\rho = \frac{101300 \times 0.029}{8.314 \times 293}$$

$$\rho = 1.2059\ \mathrm{kg\,m^{-3}} \tag{12}$$

The Chilton-Colburn relation converts the measured mass coefficient to a heat coefficient:

$$h = \rho C_p k_c \left(\frac{\mathrm{Sc}}{\mathrm{Pr}}\right)^{2/3}$$

Substituting the air properties gives

$$h = 1.2059(1007)(0.1177)\left(\frac{1.0333}{0.71}\right)^{2/3} = 183.55\ \mathrm{W\,m^{-2}\,K^{-1}}$$

The surface fraction is about 0.0468, so the low-flux approximation is plausible but not exact.

External heating maintains the specified surface temperature. The estimated heat coefficient describes how this flow would respond to a small temperature difference; it does not supply latent heat when the air and surface are isothermal.

Answer

The concentration-basis mass transfer coefficient is $k_c \simeq 0.118\ \mathrm{m\,s^{-1}}$. The corresponding heat transfer coefficient is $h \simeq 184\ \mathrm{W\,m^{-2}\,K^{-1}}$.

Back to Exercise 1

Exercise 2. Naphthalene sheet

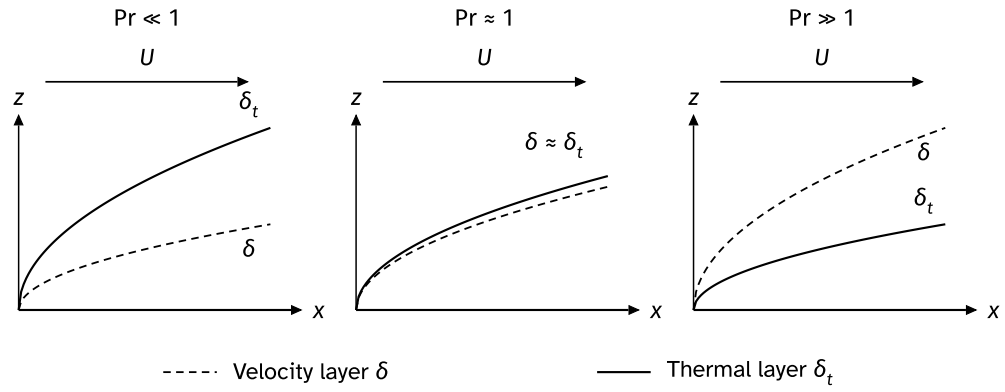


Figure 3: Air flows along the naphthalene sheet. The velocity boundary layer is the same type as for a heated plate; the concentration boundary layer replaces the thermal boundary layer in the mass transfer analogy.

Use the concentration boundary layer in Figure 2, with the leading edge at the start of the sheet. The one-metre dimension is along the flow; width remains unspecified.

Data

- Sheet length $L = 1\ \mathrm{m}$
- Air speed $U = 20\ \mathrm{m\,s^{-1}}$
- Film diffusivity $D_{AB} = 6 \times 10^{-6}\ \mathrm{m^2\,s^{-1}}$
- Film kinematic viscosity $\nu = 1.5 \times 10^{-5}\ \mathrm{m^2\,s^{-1}}$
- Surface concentration

$$c_{A,s} = 10^{-5}\ \mathrm{kmol\,m^{-3}} = 0.0100\ \mathrm{mol\,m^{-3}}$$

Assumptions

- Steady transfer from a uniform saturated surface.
- Negligible naphthalene in the incoming air.
- Constant properties and negligible normal injection.
- The source estimate treats the boundary layer as turbulent from the leading edge. A separate estimate allows natural transition.

Calculations

Use length-based flow and transfer groups:

$$\mathrm{Re}_L = \frac{UL}{\nu}, \qquad \mathrm{Sc} = \frac{\nu}{D_{AB}}$$

The groups are

$$\mathrm{Re}_L = \frac{20}{1.5 \times 10^{-5}} = 1.3333 \times 10^{6}$$

and $\mathrm{Sc} = 2.5$.

For the source's fully turbulent model, first determine the average Sherwood number, then recover the coefficient from

$$\bar{k}_c = \frac{\overline{\mathrm{Sh}}_L D_{AB}}{L}.$$

The transfer area per unit width is L, so the loss per unit width is

$$\frac{\dot{n}_A}{b} = \bar{k}_c L (c_{A,s} - c_{A,\infty}).$$

Substitution gives

$$\overline{\mathrm{Sh}}_L = 0.0366 \mathrm{Re}_L^{0.8} \mathrm{Sc}^{1/3}$$

$$\overline{\mathrm{Sh}}_L = 3945.3$$

Use the same one-metre length in the dimensional conversion:

$$\bar{k}_c = \frac{\overline{\mathrm{Sh}}_L D_{AB}}{L} = \frac{3945.3(6 \times 10^{-6})}{1.0} = 0.02367\ \mathrm{m\,s^{-1}}$$

$$\frac{\dot{n}_A}{\text{width}} = \bar{k}_c L c_{A,s}$$

$$\frac{\dot{n}_A}{b} = 2.367 \times 10^{-4}\ \mathrm{mol\,s^{-1}\,m^{-1}} \tag{13}$$

Here b denotes sheet width in metres.

The table uses 0.037, which changes this estimate by about one percent.

For an untripped plate, the laminar starting length is

$$x_{tr} = \frac{5 \times 10^5 \nu}{U} = 0.375\,\mathrm{m}.$$

The mixed average relation

$$\overline{\mathrm{Sh}}_L = (0.037\mathrm{Re}_L^{0.8} - 871)\mathrm{Sc}^{1/3}$$

gives 2806.4, $\bar{k}_c = 0.01684$ m s⁻¹ and $\dot{n}_A/\text{width} = 1.684 \times 10^{-4}$ mol s⁻¹ m⁻¹.

This is the preferred estimate when transition occurs naturally at the stated Reynolds number.

Both answers are conditional on the specified flow model; they are not interchangeable rounding choices.

With $M_A = 0.128$ kg mol⁻¹ the source fully turbulent estimate corresponds to 2.62 kg day⁻¹ per metre of width.

Answer

With the source's fully turbulent model, $\bar{k}_c = 0.0237\ \mathrm{m\,s^{-1}}$ and the loss per unit width is $2.37 \times 10^{-4}\ \mathrm{mol\,s^{-1}\,m^{-1}}$. For an untripped plate with natural transition, the preferred estimates are $0.0168\ \mathrm{m\,s^{-1}}$ and $1.68 \times 10^{-4}\ \mathrm{mol\,s^{-1}\,m^{-1}}$.

Back to Exercise 2

Exercise 3. Wet-bulb thermometer

Data

Quantity	Value
Wet-bulb temperature	$T_s = 290\ \mathrm{K}$
Surface vapour pressure	$p_{A,s} = 1940\ \mathrm{Pa}$
Air density	$\rho = 1.219\ \mathrm{kg\,m^{-3}}$
Specific heat capacity	$C_p = 1006\ \mathrm{J\,kg^{-1}\,K^{-1}}$
Prandtl number	$\mathrm{Pr} = 0.71$
Schmidt number	$\mathrm{Sc} = 0.61$
Latent heat	$\lambda = 2.461 \times 10^6\ \mathrm{J\,kg^{-1}}$
Water molar mass	$M_A = 0.018\ \mathrm{kg\,mol^{-1}}$

Assumptions

- Steady wick temperature and evaporation rate.
- Dry incoming air and saturated gas at the wet surface.
- Radiation and conduction through the thermometer stem are neglected.
- Sensible energy transport by injected vapour and blowing are neglected.
- Constant properties and matching heat and mass transfer conditions.

Calculations

At steady state convection supplies evaporation's latent heat.

With zero incoming vapour,

$$h(T_\infty - T_s) = \lambda M_A k_c c_{A,s}$$

$$\frac{k_c}{h} = \frac{1}{\rho C_p}\left(\frac{\mathrm{Pr}}{\mathrm{Sc}}\right)^{2/3} \tag{14}$$

The ratio is $9.023 \times 10^{-4}\ \mathrm{m^3\ K/J}$, and

$$c_{A,s} = \frac{1940}{8.314 \times 290} = 0.80465\ \mathrm{mol\,m^{-3}}.$$

The energy balance gives the temperature rise above the wet bulb:

$$T_\infty - T_s = \lambda M_A c_{A,s} \frac{k_c}{h}$$

Substituting the latent heat, concentration and coefficient ratio gives

$$T_\infty - T_s = 2.461 \times 10^6 (0.018)(0.80465)(9.023 \times 10^{-4}) = 32.161\ \mathrm{K}$$

Adding this temperature difference to the wet-bulb reading gives

$$T_\infty = 290 + 32.161 = 322.16\ \mathrm{K}$$

Velocity cancels because the same flow affects both coefficients.

This cancellation holds under the stated analogy and energy assumptions; it does not remove ventilation requirements or radiative errors of a real thermometer.

Answer

The estimated dry-air temperature is $T_\infty = 322.2\ \mathrm{K}$, approximately $49.0\,^\circ\mathrm{C}$.

Back to Exercise 3

Exercise 4. Heat and mass analogy

Data

Quantity	Value
Air speed	$U = 15\ \mathrm{m\,s^{-1}}$
Wet area	$A_s = 1.5\ \mathrm{m^2}$
Supplied heat correlation	$\bar{h} = 21U^{0.6}$
Air density	$\rho = 1.128\ \mathrm{kg\,m^{-3}}$
Specific heat capacity	$C_p = 1007\ \mathrm{J\,kg^{-1}\,K^{-1}}$
Kinematic viscosity	$\nu = 15.89 \times 10^{-6}\ \mathrm{m^2\,s^{-1}}$
Diffusivity	$D_{AB} = 2.6 \times 10^{-5}\ \mathrm{m^2\,s^{-1}}$
Prandtl number	$\mathrm{Pr} = 0.71$
Surface saturation pressure	$p_A^{\mathrm{sat}} = 5624\ \mathrm{Pa}$
Surface temperature	$T_s = 308.15\ \mathrm{K}$
Water molar mass	$M_A = 0.018\ \mathrm{kg\,mol^{-1}}$

The transport properties are rounded model inputs. The surface saturation pressure is consistent with 35 degrees Celsius using the NIST Antoine correlation.

Assumptions

- Steady transfer from a uniformly wet surface.
- Dry incoming air.
- The surface is maintained at the stated air temperature.
- Constant supplied properties.
- Heat and mass transfer share the same flow and surface conditions.
- Normal evaporation flow is small.

Calculations

The measurement gives

$$\bar{h} = 21 \times 15^{0.6} = 106.63\,\mathrm{W\,m^{-2}\,K^{-1}}.$$

Compute $\mathrm{Sc} = 0.61115$.

The matching-flow Chilton-Colburn relation is

$$\bar{k}_c = \frac{\bar{h}}{\rho C_p}\left(\frac{\mathrm{Pr}}{\mathrm{Sc}}\right)^{2/3}$$

Substituting the coefficient and property values gives

$$\begin{aligned} \bar{k}_c &= \frac{106.63}{1.128 \times 1007}\left(\frac{0.71}{0.61115}\right)^{2/3} \\ \bar{k}_c &= 0.10374\,\mathrm{m\,s^{-1}} \end{aligned} \tag{15}$$

At 308.15 K,

$$c_{A,s} = \frac{5624}{8.314 \times 308.15} = 2.1952\,\mathrm{mol\,m^{-3}}.$$

The evaporation rate is coefficient times area times concentration difference:

$$\dot{n}_A = \bar{k}_c A_s (c_{A,s} - c_{A,\infty})$$

With dry bulk air, substitution gives

$$\dot{n}_A = 0.10374(1.5)(2.1952) = 0.3416\,\mathrm{mol\,s^{-1}}$$

Convert the molar rate to mass rate:

$$\dot{m}_A = M_A \dot{n}_A = 0.018(0.3416) = 0.00615\,\mathrm{kg\,s^{-1}} = 22.1\,\mathrm{kg\,h^{-1}}$$

Dryness and maintained surface temperature are explicit additions to the source question.

At 50 percent relative humidity with equal surface and ambient temperature the driving force and rate halve.

The saturated surface fraction is about 0.0555, exceeding the source's 0.05 rule of thumb; a blowing correction may matter.

Maintaining the warm surface requires heat input.

This is not an adiabatic wet-bulb calculation.

Answer

Using the supplied additional properties, $\bar{k}_c \simeq 0.104\,\mathrm{m\,s^{-1}}$ and

$$\dot{m}_A \simeq 0.00615\,\mathrm{kg\,s^{-1}} = 22.1\,\mathrm{kg\,h^{-1}}.$$

These estimates require the stated dry-air and maintained-surface-temperature assumptions.

Back to Exercise 4

Exercise 5. Annular transfer

Data

- Inner diameter d_1
- Outer diameter d_2
- Reynolds number based on inner diameter Re_{d_1}
- Empirical correlation constant C

No numerical geometry, velocity or value of the correlation constant is supplied. Here the symbol for the correlation constant does not denote concentration.

Assumptions

- The heat and mass problems have matching geometry and flow regime.
- Uniform surface concentration corresponds to the heat correlation's wall condition.
- Low mass flux leaves the velocity field essentially unchanged.
- The gas is dilute and isothermal, so wall and bulk viscosities are approximately equal.
- The translated correlation is used within the applicable Reynolds and Schmidt ranges.

Calculations

Keep the same geometry, Reynolds definition and experimental range.

Replace Nusselt by Sherwood and Prandtl by Schmidt.

At isothermal dilute conditions $\mu_s \approx \mu$, giving

$$\mathrm{Sh}_{d_1} = C\mathrm{Re}_{d_1}^{0.8}\mathrm{Sc}^{1/3}\left(\frac{d_2}{d_1}\right)^{0.15}$$

$$k_c = \frac{D_{AB}}{d_1}\mathrm{Sh}_{d_1} \tag{17}$$

The viscosity factor can be retained if wall temperature or composition changes viscosity appreciably, but the heat correlation alone does not validate a mass-composition correction.

The Reynolds and diameter-ratio factors describe flow and geometry, so they remain unchanged. The viscosity correction becomes unity under the stated isothermal dilute approximation.

Answer

For the stated isothermal, dilute model,

$$k_c = \frac{D_{AB}}{d_1}C\mathrm{Re}_{d_1}^{0.8}\mathrm{Sc}^{1/3}\left(\frac{d_2}{d_1}\right)^{0.15}$$

The answer remains symbolic because the required numerical data are not supplied.

Back to Exercise 5

Exercise 6. Wet felt pipe

The inlet, outlet and wetted wall are shown in Figure 4. Concentration rises along the pipe, so the downstream driving force is smaller than the inlet driving force.

Data

Quantity	Value
Pipe diameter	$d = 0.15\ \mathrm{m}$
Pipe length	$L = 6\ \mathrm{m}$
Bulk air speed	$U = 1.5\ \mathrm{m\,s^{-1}}$
Dry inlet temperature	$T_i = 310\ \mathrm{K}$
Wet wall temperature	$T_s = 290\ \mathrm{K}$
Surface vapour pressure	$p_{A,s} = 1917\ \mathrm{Pa}$
Kinematic viscosity	$\nu = 1.569 \times 10^{-5}\ \mathrm{m^2\,s^{-1}}$
Diffusivity	$D_{AB} = 2.626 \times 10^{-5}\ \mathrm{m^2\,s^{-1}}$
Fanning friction factor	$C_f = 0.007$
Water molar mass	$M_A = 0.018\ \mathrm{kg\,mol^{-1}}$

The vapour pressure, transport properties and friction factor come from the lecturer's notes.

Assumptions

- Steady flow and evaporation.
- Constant bulk velocity and supplied properties.
- Uniform saturated wall concentration.
- Each cross-section has a representative mixed bulk concentration.
- Axial diffusion and normal-flow effects are neglected.
- A developed-flow average coefficient approximates the pipe transfer.

Calculations

The surface concentration is

$$c_{A,s} = \frac{1917}{8.314 \times 290} = 0.79508\ \mathrm{mol\,m^{-3}},$$

and $c_{A,i} = 0$.

The groups are

$$\mathrm{Re}_d = \frac{1.5 \times 0.15}{\nu} = 14340$$

and $\mathrm{Sc} = 0.59749$.

The pipe spans 40 diameters; the roughly ten-diameter turbulent development estimate motivates a developed average approximation, not exact development at the inlet.

The concentration model treats the maintained wall and gas temperatures through the supplied surface concentration and constant properties. It does not solve the axial thermal field. The friction analogy gives the coefficient as

$$k_c = \frac{C_f U}{2\mathrm{Sc}^{2/3}} = 0.0074007\ \mathrm{m\,s^{-1}} \tag{19}$$

The source table's Dittus-Boelter-type mass row gives about 0.00688 m s⁻¹, but $\mathrm{Sc} = 0.5975$ falls below its stated 0.7 limit.

That comparison is an extrapolation and is not the governing answer.

The primary friction estimate also sits at the approximate lower edge of the source's Chilton-Colburn range; treat the result as an engineering estimate.

For a slice dx, steady balance is inflow minus outflow plus wall input equals zero.

Inflow is $UA_c c_A(x)$, outflow is $UA_c c_A(x+dx)$, and wall input is

$$k_c \pi d\, dx (c_{A,s} - c_A),$$

where $A_c = \pi d^2/4$.

Dividing and taking the limit yields

$$\frac{Ud}{4}\frac{dc_A}{dx} = k_c(c_{A,s} - c_A) \tag{20}$$

Integrate from inlet to exit.

For this circular pipe the Number of transfer units is

$$\mathcal{N} = \frac{4k_c L}{Ud} = \frac{k_c A_s}{Q},$$

using $A_s = \pi d L$ and $Q = U\pi d^2/4$. Small $\mathcal{N}$ leaves the gas close to its inlet composition; large $\mathcal{N}$ brings it towards the maintained wall value. For $k_c = 0.01\ \mathrm{m\,s^{-1}}$, $L = 1\ \mathrm{m}$, $U = 2\ \mathrm{m\,s^{-1}}$ and $d = 0.02\ \mathrm{m}$, $\mathcal{N} = 1$ and the remaining concentration difference is $e^{-1} = 0.368$ of its inlet value. Omitting the factor four would underpredict the contact effect.

Then

$$\begin{aligned} \frac{c_{A,s} - c_{A,e}}{c_{A,s} - c_{A,i}} &= e^{-\mathcal{N}} \\ \mathcal{N} &= 0.78941 \\ c_{A,e} &= 0.4340\ \mathrm{mol\,m^{-3}} \end{aligned} \tag{21}$$

For the dry inlet, the integrated balance can be evaluated directly:

$$\mathcal{N} = \frac{4(0.0074007)(6)}{1.5(0.15)} = 0.78941$$

$$c_{A,e} = c_{A,s} - (c_{A,s} - c_{A,i})e^{-\mathcal{N}} = 0.79508(1 - e^{-0.78941}) = 0.4340 \text{ mol m}^{-3}$$

The exit is about 54.6 percent of surface saturation on this concentration basis.

Let $\Delta c_i = c_{A,s} - c_{A,i}$ and $\Delta c_e = c_{A,s} - c_{A,e}$.

The logarithmic mean is

$$\Delta c_{lm} = \frac{\Delta c_i - \Delta c_e}{\ln\left(\frac{\Delta c_i}{\Delta c_e}\right)} \tag{22}$$

$$\Delta c_{lm} = 0.54981 \text{ mol m}^{-3}$$

It is the exact mean driving force for this exponential model.

At equal end differences its limiting value is that common difference.

Apply the mean driving force over wall area πdL:

$$\dot{n}_A = k_c \pi dL \Delta c_{lm} = 0.0074007\pi(0.15)(6)(0.54981) = 0.01150 \text{ mol s}^{-1}$$

An independent overall balance gives

$$\dot{n}_A = UA_c(c_{A,e} - c_{A,i}) = 1.5\frac{\pi(0.15)^2}{4}(0.4340 - 0) = 0.01150 \text{ mol s}^{-1}$$

Agreement confirms that the wall-transfer calculation and outlet concentration satisfy the same material balance.

Thus the final replenishment rate is

$$\dot{m}_A = 0.018\dot{n}_A = 2.07 \times 10^{-4} \text{ kg s}^{-1},$$

or 0.745 kg h^{-1}.

Answer

The required water replenishment rate is approximately 2.07×10^{-4} kg s^{-1}, or 0.745 kg h^{-1}. The exit bulk concentration is 0.434 mol m^{-3}.

Back to Exercise 6

Exercise 7. Warm water pan

Data

Quantity	Value
Pan diameter	$d = 0.225$ m
Rounded surface temperature	$T_s = 310$ K
Rounded air temperature	$T_\infty = 290$ K
Pressure	$P = 101325$ Pa

Quantity	Value
Gravitational acceleration	$g = 9.81\ \mathrm{m\,s^{-2}}$
Kinematic viscosity	$\nu = 15.89 \times 10^{-6}\ \mathrm{m^2\,s^{-1}}$
Diffusivity	$D_{AB} = 2.6 \times 10^{-5}\ \mathrm{m^2\,s^{-1}}$
Prandtl number	$\mathrm{Pr} = 0.707$
Specific heat capacity	$C_p = 1006\ \mathrm{J\,kg^{-1}\,K^{-1}}$
Air conductivity	$k = 0.0263\ \mathrm{W\,m^{-1}\,K^{-1}}$
Film density	$\rho = 1.1769\ \mathrm{kg\,m^{-3}}$
Expansion coefficient	$\beta = 1/300\ \mathrm{K^{-1}}$
Surface saturation pressure	$p_A^{\mathrm{sat}} = 6220\ \mathrm{Pa}$
Water molar mass	$M_A = 0.018\ \mathrm{kg\,mol^{-1}}$
Air molar mass	$M_B = 0.029\ \mathrm{kg\,mol^{-1}}$

Use the source's rounded temperatures and the supplementary properties from the lecturer's notes.

Assumptions

- Steady natural convection above an upward-facing wet surface.
- Dry bulk air and saturated gas at the water surface.
- Uniform maintained surface temperature.
- Constant supplied film properties and ideal gas behaviour.
- Low normal evaporation flux.
- Composition buoyancy and thermal buoyancy are considered separately, as requested.

Calculations

Area is

$$A_s = \frac{\pi(0.225)^2}{4} = 0.039761\ \mathrm{m^2}$$

and horizontal characteristic length is

$$L = \frac{A_s}{P_s} = \frac{d}{4} = 0.05625\ \mathrm{m}.$$

Using diameter instead gives a Grashof number 64 times too large.

The surface water mass concentration is

$$\rho_{A,s} = \frac{p_{A,s} M_A}{R T_s} = 0.043440\ \mathrm{kg\,m^{-3}}$$

and bulk water concentration is zero.

The Schmidt number is 0.61115.

For composition alone compare dry and saturated air at the same 310 K.

The densities are

$$\rho_{B,310} = \frac{PM_B}{RT_s} = 1.1401$$

and

$$\rho_{s,310} = \rho_{A,s} + \rho_{B,s}$$

$$\rho_{A,s} = \frac{6220(0.018)}{8.314(310)}$$

$$\rho_{B,s} = \frac{(101325 - 6220)(0.029)}{8.314(310)}$$

$$\rho_{s,310} = 1.1136\ \mathrm{kg\,m^{-3}} \tag{23}$$

Thus $\Delta\rho_D = 0.02655$ kg m^{-3}.

Use their mean density 1.12685 kg m^{-3} in Eq. 18.

The composition Grashof number is

$$\mathrm{Gr}_D = \frac{g\Delta\rho_D L^3}{\rho\nu^2}$$

Substituting the same-temperature density difference gives

$$\mathrm{Gr}_D = \frac{9.81(0.02655)(0.05625)^3}{1.12685(15.89\times10^{-6})^2} = 1.6291\times10^5$$

The composition Rayleigh number is $\mathrm{Ra}_D = \mathrm{Gr}_D\mathrm{Sc} = 9.9562\times10^4$, within the specified horizontal-plate range.

The correlation gives

$$\mathrm{Sh} = 0.54\mathrm{Ra}_D^{1/4} = 0.54(9.9562\times10^4)^{1/4} = 9.5922$$

$$k_c = \frac{\mathrm{Sh}D_{AB}}{L} = \frac{9.5922(2.6\times10^{-5})}{0.05625} = 0.0044337\ \mathrm{m\,s^{-1}}$$

$$\dot{m}_A = k_c A_s \rho_{A,s} = 0.0044337(0.039761)(0.043440)$$

$$\dot{m}_A = 7.6580\times10^{-6}\ \mathrm{kg\,s^{-1}} \tag{24}$$

The composition-only estimate is 27.6 g h^{-1}.

Under the matching-flow transfer analogy, the corresponding heat coefficient is

$$h = \rho C_p k_c \left(\frac{\mathrm{Sc}}{\mathrm{Pr}}\right)^{2/3} = 1.1769(1006)(0.0044337)\left(\frac{0.61115}{0.707}\right)^{2/3} \simeq 4.76\ \mathrm{W\,m^{-2}\,K^{-1}}$$

For thermal buoyancy alone, use the thermal form of Eq. 18:

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

Substitution gives

$$\mathrm{Gr}_T = \frac{9.81\left(\frac{1}{300}\right)(310 - 290)(0.05625)^3}{(15.89 \times 10^{-6})^2} = 4.6100 \times 10^5$$

The thermal Rayleigh number is $\mathrm{Ra}_T = \mathrm{Gr}_T\mathrm{Pr} = 3.2593 \times 10^5$. The matching heat correlation gives

$$\mathrm{Nu} = 0.54\mathrm{Ra}_T^{1/4} = 0.54(3.2593 \times 10^5)^{1/4} = 12.903$$

Recover the heat transfer coefficient using the thermal conductivity:

$$h = \frac{\mathrm{Nu}k}{L} = \frac{12.903(0.0263)}{0.05625} = 6.033\ \mathrm{W\,m^{-2}\,K^{-1}}$$

For the thermally driven flow, convert h to a mass coefficient:

$$k_c = \frac{h}{\rho C_p}\left(\frac{\mathrm{Pr}}{\mathrm{Sc}}\right)^{2/3} = \frac{6.033}{1.1769(1006)}\left(\frac{0.707}{0.61115}\right)^{2/3} = 0.005615\ \mathrm{m\,s^{-1}}$$

The evaporation rate then follows from the surface-to-bulk mass-concentration difference:

$$\begin{aligned}\dot{m}_A &= k_c A_s(\rho_{A,s} - \rho_{A,\infty}) \\ &= 0.005615(0.039761)(0.043440) \\ &= 9.698 \times 10^{-6}\ \mathrm{kg\,s^{-1}} = 34.91\ \mathrm{g\,h^{-1}}\end{aligned}$$

These are the two requested isolated-mechanism answers.

For comparison, using dry-air density at 290 K and saturated-air density at 310 K gives a total density difference that includes both mechanisms.

The mean density 1.16616 and total difference 0.10515 kg m⁻³ give $\mathrm{Gr} = 6.2366 \times 10^5$, $\mathrm{Sh} = 13.419$ and $k_c = 0.006202$ m s⁻¹.

The corresponding total-density estimate is $\dot{m}_A = 1.0712 \times 10^{-5}$ kg s⁻¹, about 38.6 g h⁻¹.

This approximately 39 g h⁻¹ estimate already includes thermal expansion. Adding a thermal Grashof contribution would count that expansion twice.

The real pan has reinforcing thermal and composition buoyancy.

The total-density route is a possible single-group engineering approximation; an exact coupled answer requires a validated heat-and-species natural-convection model.

The two isolated evaporation rates cannot be added to predict the combined rate, because the same velocity field transports both heat and species.

The saturation fraction is 0.0614, so low-flux and dry-air-property assumptions remain approximate.

The maintained water temperature implies external heat supply; no latent-heat constraint has been omitted from an adiabatic balance because this pan is not adiabatic.

Answer

For composition-driven convection alone, $k_c = 0.00443\,\mathrm{m\,s^{-1}}$, $h \simeq 4.76\,\mathrm{W\,m^{-2}\,K^{-1}}$ and $\dot{m}_A \simeq 27.6\,\mathrm{g\,h^{-1}}$.

For temperature-driven convection alone, $h = 6.03\,\mathrm{W\,m^{-2}\,K^{-1}}$, $k_c = 0.00562\,\mathrm{m\,s^{-1}}$ and $\dot{m}_A \simeq 34.9\,\mathrm{g\,h^{-1}}$.

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Exercise 8. Evaporation from a tray and an outlet-humidity check

Data

Quantity	Value
Tray length	$L = 0.60\,\mathrm{m}$
Tray width	$b = 0.25\,\mathrm{m}$
Duct height	$H = 0.12\,\mathrm{m}$
Air speed	$U = 2.0\,\mathrm{m\,s^{-1}}$
Temperature	$T = 303.15\,\mathrm{K}$
Pressure	$P = 101325\,\mathrm{Pa}$
Kinematic viscosity	$\nu = 1.60 \times 10^{-5}\,\mathrm{m^2\,s^{-1}}$
Diffusivity	$D_{AB} = 2.60 \times 10^{-5}\,\mathrm{m^2\,s^{-1}}$
Surface saturation pressure	$p_A^{\mathrm{sat}} = 4240\,\mathrm{Pa}$
Water molar mass	$M_A = 0.0180\,\mathrm{kg\,mol^{-1}}$

Assumptions

- Steady external transfer with a boundary layer starting at the tray leading edge.
- Uniform incoming velocity and negligible upstream profile effects.
- Maintained saturated surface at the air temperature.
- Constant properties and dilute transfer without a Stefan-flow correction.
- Dry bulk air is used for the initial rate estimate and then checked.

Calculations

The concentration boundary layer grows along the tray, so use its length $L = 0.60$ m for external transfer. The duct hydraulic diameter is not the length in this flat-plate correlation:

$$\mathrm{Re}_L = \frac{2.0(0.60)}{1.60 \times 10^{-5}} = 75000 \qquad \mathrm{Sc} = \frac{1.60 \times 10^{-5}}{2.60 \times 10^{-5}} = 0.6154$$

The layer is laminar, and Sc exceeds 0.5. The average flat-plate correlation and coefficient definition are

$$\overline{\mathrm{Sh}}_L = 0.664 \mathrm{Re}_L^{1/2} \mathrm{Sc}^{1/3}$$

$$\bar{k}_c = \frac{\overline{\mathrm{Sh}}_L D_{AB}}{L}$$

Substitution gives

$$\overline{\mathrm{Sh}}_L = 0.664 \mathrm{Re}_L^{1/2} \mathrm{Sc}^{1/3} = 154.7$$

$$\bar{k}_c = \frac{\overline{\mathrm{Sh}}_L D_{AB}}{L} = 0.006702 \,\mathrm{m\,s^{-1}}$$

At the interface the vapour is saturated. With $T = 303.15$ K,

$$c_{As} = \frac{4240}{8.314T} = 1.682 \,\mathrm{mol\,m^{-3}}.$$

Approximate the bulk concentration as zero and use transfer area

$$A_s = 0.60(0.25) = 0.150 \,\mathrm{m^2}:$$

$$\dot{n}_A = \bar{k}_c A_s (c_{As} - c_{A,\infty}) = 0.006702(0.150)(1.682 - 0) = 1.691 \times 10^{-3} \,\mathrm{mol\,s^{-1}}$$
$$\dot{m}_A = M_A \dot{n}_A = 0.1096 \,\mathrm{kg\,h^{-1}}$$

For the outlet balance use the duct flow area,

$$A_{duct} = 0.25(0.12) = 0.0300 \,\mathrm{m^2}.$$

The dry-air molar flow is

$$\dot{n}_{air} = u A_{duct} \frac{P}{RT} = 2.412 \,\mathrm{mol\,s^{-1}}$$

The outlet contains both dry air and the added vapour:

$$y_{A,out} = \frac{\dot{n}_A}{\dot{n}_{air} + \dot{n}_A} = 7.007 \times 10^{-4}$$

$$\mathrm{RH}_{out} = \frac{y_{A,out} P}{p_A^{sat}} = 0.01674 = 1.67 \ \text{ percent}$$

Under the assumed mixed-bulk description, the bulk concentration reaches only 1.67 percent of the surface saturation concentration. The driving force therefore falls by less than 2 percent from inlet to outlet. This supports using dry bulk air in the first evaporation estimate. A much wetter outlet would require coupling local transfer to the changing bulk humidity.

This humidity check does not remove all model approximations. The surface mole fraction is

$$\frac{4240}{101325} = 0.04185,$$

so a Stefan-flow correction can also change the dilute estimate by a few percent. The imposed heater avoids an evaporative-cooling calculation; without it, the surface temperature and vapour pressure would need a coupled energy balance.

Answer

The average mass transfer coefficient is $\bar{k}_c = 0.00670\,\mathrm{m\,s^{-1}}$ and the evaporation rate is $0.110\,\mathrm{kg\,h^{-1}}$. The estimated outlet relative humidity is 1.67 percent, supporting the dry-bulk approximation.

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Exercise 9. Drying a thin coating on a spherical pellet

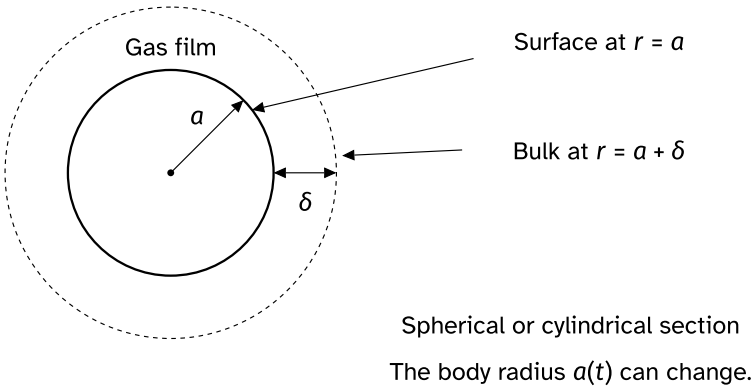


Figure 5: A thin solvent coating surrounds the pellet. Its surface concentration is fixed by vapour equilibrium, and solvent transfers into the surrounding air. The diagram shows the external gas film; the model tracks coating loss rather than diffusion inside the pellet.

Data

Quantity	Value
Pellet diameter	$D = 0.010\ \mathrm{m}$
Air speed	$u = 0.10\ \mathrm{m\,s^{-1}}$
Initial solvent loading	$m_A/A_s = 0.0200\ \mathrm{kg\,m^{-2}}$
Solvent diffusivity	$D_{AB} = 9.62 \times 10^{-6}\ \mathrm{m^2\,s^{-1}}$
Solvent saturation pressure	$p_A^{\mathrm{sat}} = 1270\ \mathrm{Pa}$
Solvent molar mass	$M_A = 0.0780\ \mathrm{kg\,mol^{-1}}$
Liquid density	$\rho_L = 876.5\ \mathrm{kg\,m^{-3}}$
Air and surface temperature	$T = T_s = 300\ \mathrm{K}$

Use the supplied air properties.

Assumptions

- Quasi steady external transfer and constant supplied properties.
- Negligible solvent in the bulk air.
- The exposed coating remains saturated until the solvent is exhausted.
- Negligible coating resistance and normal evaporation flow.
- Surface temperature is maintained by external heat supply.
- Fixed pellet diameter is used and the coating thickness is checked.

Calculations

Use pellet diameter for both the flow Reynolds number and the transfer groups:

$$\mathrm{Re}_D = \frac{\rho u D}{\mu} = \frac{1.18(0.10)(0.010)}{1.85 \times 10^{-5}} = 63.78$$

$$\mathrm{Sc} = \frac{\mu}{\rho D_{AB}} = \frac{1.85 \times 10^{-5}}{1.18(9.62 \times 10^{-6})} = 1.630$$

$$\mathrm{Nu}_D = 2 + 0.6(63.78)^{1/2}(0.707)^{1/3} = 6.269$$

$$\mathrm{Sh}_D = 2 + 0.6(63.78)^{1/2}(1.630)^{1/3} = 7.639$$

This Reynolds number is within the low-Re use described for the sphere correlation in the external forced-convection notes. The two transfer groups differ because Schmidt and Prandtl numbers differ. Their additive two is the stagnant-sphere limit and must be retained.

Recover the dimensional coefficients with the corresponding transport property:

$$h = \frac{\mathrm{Nu}_D k}{D} = 16.49\ \mathrm{W\,m^{-2}\,K^{-1}}$$

$$k_c = \frac{\mathrm{Sh}_D D_{AB}}{D} = 0.007349\ \mathrm{m\,s^{-1}}$$

Since temperature is controlled, the heat coefficient need not determine the evaporation rate. The supplied vapour pressure fixes

$$c_{As} = \frac{1270}{8.314(300)} = 0.5092\,\mathrm{mol\,m^{-3}}.$$

With dry bulk air,

$$N_A = k_c(c_{As} - c_{A,\infty}) = 0.007349(0.5092 - 0) = 0.003742\ \mathrm{mol\,m^{-2}\,s^{-1}}$$

The spherical area is

$$A_s = \pi D^2 = 3.1416 \times 10^{-4}\,\mathrm{m^2}.$$

The loading converts as $0.0020\ \mathrm{g\ cm^{-2}} = 0.0200\ \mathrm{kg\ m^{-2}}$, so the initial solvent mass is $m_{A0} = 6.283 \times 10^{-6}$ kg. The constant-rate estimate is

$$t_{dry} = \frac{m_{A0}}{M_A N_A A_s} = \frac{0.0200}{0.0780(0.003742)} = 68.5\ \mathrm{s}$$

The area cancels because both inventory and evaporation rate scale with it. Diameter still affects drying time through the transfer coefficient.

The solvent-equivalent coating thickness is

$$\frac{0.0200}{876.5} = 2.28 \times 10^{-5}\ \mathrm{m},$$

only 0.46 percent of the pellet radius. Treating the diameter as fixed is consistent with a thin coating provided the nonvolatile coating is also thin. The surface mole fraction is 0.01253, supporting the dilute approximation. Actual paint can dry more slowly as solvent activity falls or diffusion through the coating becomes limiting; the calculated time uses the stated constant-activity model.

Answer

The heat transfer coefficient is $h = 16.5\ \mathrm{W\,m^{-2}\,K^{-1}}$, the mass transfer coefficient is $k_c = 0.00735\ \mathrm{m\,s^{-1}}$ and the constant-rate drying time is approximately 68.5 s. The solvent-equivalent thickness is $22.8\ \mu\mathrm{m}$, only 0.46 percent of the pellet radius.

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