

Introduction to Mass Transfer

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1 Introduction

This document was produced and published by Gary Slenders – Principal Process Engineer for AOMC Pty Ltd as part of the education series for Engineering Student in the field of Process, Chemical, and Environmental Engineers as a means of providing a better understanding of the fundamental processes involved in Odour Control. The information contained in this document is based on established material currently in the public domain. You are free to use this material, we only ask that you provide appropriate recognition to the author if you use a significant portion of this document for educational or commercial purposes.

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The key purpose of this document is to simplify the understanding of what mass and heat transfer is all about. As such the information presented is not as rigorous presentation as seen in the standard texts. For a more standardised assessment, we always recommend going back to the source documents.

2 How concentrated mass behaves

The basis of odour control is the movement of molecules from air into water reversing the process that generates the odour in the first place.

The laws that govern this are the laws of thermodynamics and chemistry.

The simple rules are:

1. Molecules will migrate from points of high concentration to points of low concentration.
2. Rate at which they do so is a function of the driving force that is expressed in terms of the difference between high concentration & low concentration divided by the distance between the two points

3. It is possible through good design to improve the conditions by which the driving force operates in three ways:
 - a. Minimise the low concentration to a value as low as possible by the use of chemistry and biochemistry
 - b. Minimise the distance the odour molecules need to travel by the application of turbulence.
 - c. Improve the speed at which molecules can move by the application of heat.

3 Simple Model of Molecule Diffusion

The simple model of molecular diffusion is expressed as Fick's Law.

$$\text{Rate Mass Transfer} = -D \times \frac{\text{Difference in Concentration}}{\text{Distance}}$$

This can be expressed mathematically:

$$\frac{dC}{dt} = -D \frac{\partial^2 C}{\partial x^2}$$

Referred to as Fick's equation. By the addition of a volume element $\Delta V = \Delta X \times \Delta Y \times \Delta Z$

$$\Delta V \frac{dC}{dt} = -D dA \frac{\partial^2 C}{\partial x^2} \Delta x$$

Area dA is made up of element distance $\Delta Y \times \Delta Z$. Integrating the R.H.S.

$$QdC = -DdA \frac{\Delta C}{\Delta x}$$

Change of location of unit volume represents Volumetric Flow and is in the direction of Y . Concentration change is in the direction of X

$$QdC = -\frac{D}{\Delta x} (C - C_0) dA$$

¹ Units. We have a list of variables and details of units in Section 11. We have only specified S.I. units because 1 Newton = 1 kg m/s² makes sense and 1 lb_f = g_c x 1 lb x 1 ft/s² where g_c=32.174 lb_f .s²/lb.ft makes no sense!

If you let

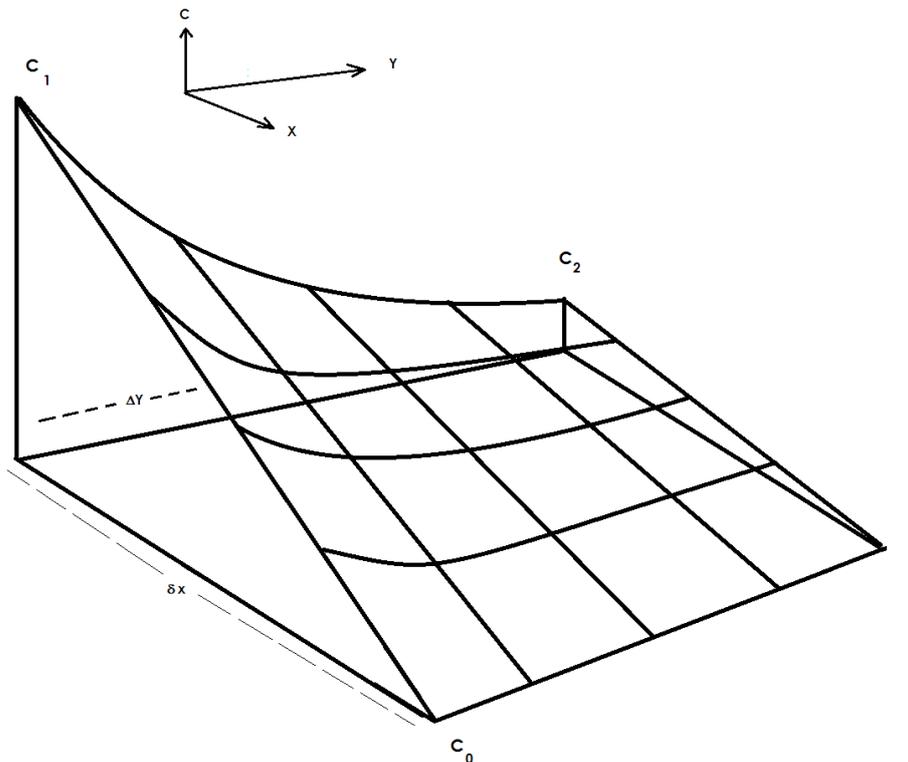
$$\frac{D}{\Delta x} = k_c,$$

k_c is the mass transfer coefficient (temperature dependent)

To finalise the equation:

$$Q \int_{C_1}^{C_2} \frac{dC}{(C - C_0)} = -k_c \int_0^A dA$$

$$\ln \left(\frac{(C_2 - C_0)}{(C_1 - C_0)} \right) = -\frac{k_c A}{Q}$$



In this simple model as viewed in the x,y,z coordinates where:

- x direction is the direction between which molecules flow between source of high concentration to a source of low concentration. Surface that adsorbs contaminant molecules.
- y direction is the direction of bulk flow of the gas
- z direction is scalar representing the concentration of contaminant C

It can be seen that the concentration of the molecular species as it progresses along the direction of flow, y direction, results in exponential decline toward a final value of C_0 . Note that if $C_0 > C_1$ then this is graph inverts to an exponential decline to higher value as molecules migrate into the air stream.

Note that temperature variability has not been highlighted. Diffusion coefficient, density, and thermal conductivity are temperature dependent.

There are limitations to this model that restricts its application. The main limitation is that there is an assumption that the carrier fluid only ever moves in one single direction and never moves laterally (in the x direction) at any stage.

The effects of turbulence will cause rotational eddies that will contain an x-direction component. This behaviour is chaotic and cannot be described in simple details, but can be determined statistically and empirically. There remains a transitional area just near the surface, an interface point where conditions are as described and the value of Δx is microscopically small. For the rest of it the actions of turbulence results in mixing of molecules in the bulk fluid causing an undetermined concentration gradient.

What has been described to date is a system whereby a fluid is in contact with a solid that at the surface of the solids there is a difference of concentration at the surface of the solid as compared to fluid that flows in a laminar fashion. In all practical mass transfer operations, there is not real basis for this existence.

In most cases, what we are considering is the transportation of molecules across phase interface, molecular transport across a gas / liquid interface boundary

One example that can be cited as a reasonable approximation is where gases flow through a packed bed of activated carbon where due to the presence of macroscopic and microscopic pores, contaminants in contact with the surface cause those contaminants to condense into a stable liquid inside the microscopic pores. This results in a solid contact surface that closely approximate a zero-concentration surface. The macroscopic pores act as a fine passage for the gas to flow and the passage is almost laminar in characteristics at that level.

By using the diffusion coefficient and estimating a value of the diffusion path to the AC surface it is possible to apply the design formula. This remains true up until a point where the microscopic pores are filled. After this point an equilibrium is established and mass transfer stops as a result of zero driving force.

4 Fourier's Equation

Although we are concerned over mass transfer, the operation of other transport phenomenon does demonstrate that there is consistency to the theory of how motion and energy operate at the microscopic level.

Apart from heat and temperature, there is also momentum and flow resistance.

The simple model of Thermal diffusion is expressed as Fourier's Law.

$$\text{Rate Heat Transfer} = -k \times \frac{\text{Difference in Temperature}}{\text{Distance}}$$

This can be expressed mathematically:

$$\rho C_p \frac{dT}{dt} = -k \frac{\partial^2 T}{\partial x^2}$$

Where ρ is the density of the fluid (important to note that in this application it is a constant) and C_p is the thermal conductivity of the fluid. By the addition of a volume element $\Delta V = \Delta X \times \Delta Y \times \Delta Z$

$$\Delta V \rho C_p \frac{dT}{dt} = -k dA \frac{\partial^2 T}{\partial x^2} \Delta x$$

Area dA is made up of element distance $\Delta Y \times \Delta Z$. Integrating the R.H.S.

$$Q \rho C_p dT = -kA \frac{\Delta T}{\Delta x} \text{ or } dE = -k dA \frac{\Delta T}{\Delta x}$$

Where E represents the thermal energy of the Change of location of unit volume represents Volumetric Flow and is in the direction of Y . Temperature change is in the direction of X

$$Q \rho C_p dT = -\frac{k}{\Delta x} (T - T_0) dA$$

If you let

$$\frac{k}{\Delta x} = h$$

where h is the heat transfer coefficient

$$Q\rho Cp \int_{T_1}^{T_2} \frac{dT}{(T - T_0)} = -h \int_0^A dA$$

$$\ln\left(\frac{(T_2 - T_0)}{(T_1 - T_0)}\right) = -\frac{hA}{Q\rho Cp}$$

This is evidence that the processes of heat transfer and mass have the same mechanism under conditions of laminar flow.

Another special case for consideration in thermal heat transfer is the situation in which you have a stationary solid plate with temperature T_1 on one side and T_2 on the other.

$$dE = -\frac{k}{\Delta x}(T - T_0)dA$$

$$E = -\frac{k}{\Delta x}(T_1 - T_2)A$$

This is the equation for conduction across a solid surface where E is the power supplied/received as the result of that temperature difference. It is the exact equation to which thermal insulation is designed for, and it is the case where it is used in combination with convective heat transfer for which this is the equation across the length of pipe.

5 Momentum Analogy

The simple model of momentum diffusion is expressed as Newton's Law.

$$\text{Rate Momentum Transfer} = -\mu \times \frac{\text{Difference in velocity}}{\text{Distance}}$$

This can be expressed mathematically:

$$\rho \frac{dv}{dt} = -\mu \frac{\partial^2 v}{\partial x^2}$$

Where ρ is the density of the fluid (important to note that in this application it is a constant). By the addition of a volume element $\Delta V = \Delta X \times \Delta Y \times \Delta Z$

$$\Delta V \rho \frac{dv}{dt} = -\mu A \frac{\partial^2 v}{\partial x^2} dx$$

Area A is made up of element distance $\Delta Y \times \Delta Z$. Integrating the R.H.S.

$$\frac{m}{A} \frac{dv}{dt} = -\mu \frac{\Delta v}{\Delta x} \text{ or } \tau = -\mu \frac{\Delta v}{\Delta x}$$

Where τ represents the shear-stress applied to a fluid and the momentum change due to the internal resistance related to viscosity

6 The limitations of Fick's, Fourier's and Newton's analogy.

The limitations of Fick's Fourier's and Newton's analogies is that due to the perpendicular velocity induced by turbulence they very rarely apply to real world situations. Bulk flow in the direction of the transport direction immediately alters randomly the value of ΔX as each point. The chaotic nature of the flow cannot be accounted for in these equations.

The only stage at which these equations are obeyed is at the laminar boundary on a stationary surface.

The other key point is that mass transfer and heat transfer does not occur of just a fluid phase in contact with a solid phase. The reality for mass transfer is that there is liquid phase spread over an extended surface in contact with gas passing across the same surface. Transfer occurs at the interface between the liquid and gas.

The design concept in the equations is that of a theoretical flat plate area. It does not have any tie in with the extended surface area design that is utilised in a packed column or a heat exchanger.

Addressing the issue of turbulence is grouped into a number of theoretical assessments:

1. Accounting the aspects of momentum and turbulent flows.
2. Looking at the scalar factors that influence mass and heat transfer coefficients.
3. Accounting for the impact of having gases pass in intermittent contact with liquids and what the interface effect that have an influence.
4. Accounting for the impact of having a column of random packed media in place of segment of flat plate.

Although the purpose of this paper is only to introduce the concept of mass and heat transfer and not strictly consider real world situations (this will be discussed more thoroughly in the next papers). It is possible to touch onto the concept of mass and heat transfer under turbulent flow conditions.

7 How to assess turbulent flow – friction factor

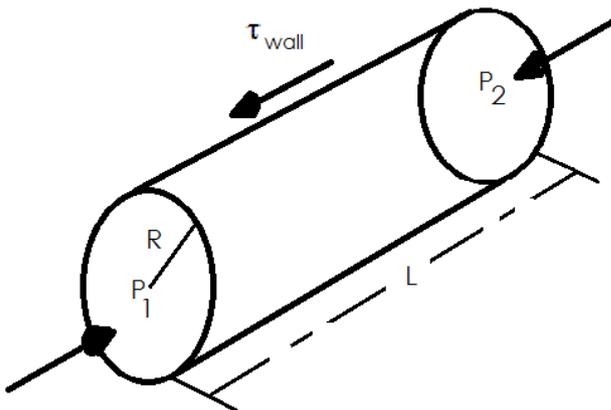
In turbulent flow conditions, all assessment work has been undertaken using laboratory observations for which empirical formula have been developed.

The main factor that described the impact of turbulence is the friction factor. The effects of applied pressure, stationary surface resistance, and resulting fluid velocity.

Most texts discuss Fanning's and Darcy's friction factors.

$$f = f_{fannings} = \frac{f_{darcy}}{4} \text{ and } f \equiv \frac{\tau_{wall}}{\rho \frac{v^2}{2}}$$

τ_{wall} = Maximum shear stress applied to wall or pipe due to flow movement.



$$(P_1 - P_2)\pi R^2 - \tau_{wall} \times 2\pi R L = 0 \text{ therefore } \tau_{wall} = \frac{(-\Delta P)R}{2L} = \rho f \frac{v^2}{2}$$

$$\frac{(-\Delta P)}{\rho} = \frac{2L}{R} f \frac{v^2}{2} = 4f \frac{L}{d} \frac{v^2}{2}$$

This is the most common expression used for calculating skin friction pressure drop in a pipe.

The method for determining the value of fanning's friction factor is by use of the modified Moody's Diagram. Note that

$$\frac{(-\Delta P)}{\rho} = 4f \frac{L}{d} \frac{v^2}{2} = f_{darcy} \frac{L}{d} \frac{v^2}{2}$$

For which Darcy friction factor is the value determined by the proper use of a Moody's Diagram.

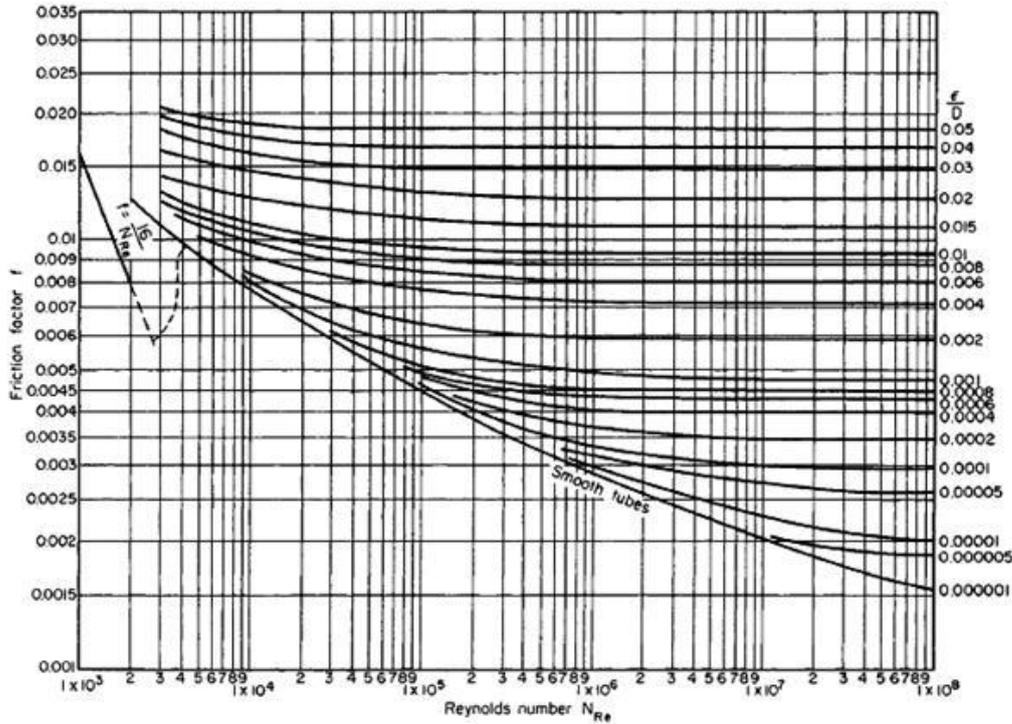


FIG. 6-9 Fanning Friction Factors. Reynolds number $Re = DV\rho/\mu$, where D = pipe diameter, V = velocity, ρ = fluid density, and μ = fluid viscosity. (Based on Moody, Trans. ASME, 66, 671 [1944].)

Reynold Number is a measure of the ratio of the flow momentum divided by the flow friction

$$N_{Re} = \frac{Dv\rho}{\mu}$$

Note that as friction factors all relate to the shear stress as applied to the wall surface, that the condition of the wall surface will have an impact on the turbulence rate. e/D represents a scalar quantity detailing the average change in in surface, the roughness, against the diameter of the pipe.

In turbulent conditions in a flowing pipe with smooth surface

$$f \approx \frac{0.054}{N_{Re}^{0.2}}$$

From this we are able to determine the mass transfer coefficient and heat transfer coefficient for the conditions of a fluid flowing inside a smooth pipe under turbulence.

8 Mass Transfer Correlation in Turbulent Conditions

Mass transfer and heat transfer under turbulent conditions has been investigated in relation to conditions of flow and correlations equations have been developed based on empirical observations.

The key factor for this is the concept of dimensionless parameters. The first as the most commonly used is that of the Reynold's Number. Dimensionless numbers simplify the parameter process as it minimises the number of variables that need to be considered.

The essential dimensionless groups in relation to turbulent mass and heat transfer are:

Dimensionless Group	Formula
Reynolds Number	$N_{Re} = \frac{dv\rho}{\mu}$
Sherwood Number	$N_{Sh} = \frac{k_c \delta x}{D}$
Schmidt Number	$N_{Sc} = \frac{\mu}{\rho D}$
Friction Factor	$f \equiv \frac{\tau_{wall}}{\rho \frac{v^2}{2}}$
Prandtl Number	$N_{Pr} = \frac{C_p \mu}{k}$
Nusselt Number	$N_{Nu} = \frac{h \delta x}{k}$
Weber Number	$N_{We} = \frac{v_l^2 \rho_l}{\sigma_l a}$
Froude Number	$N_{Fr} = \frac{v_l^2 a}{g}$

Chilton – Colburn analogy based on empirical measurements was able to determine a relationship between dimensionless factors.

In the case of mass transfer and heat transfer:

$$j_M = \frac{N_{Sh}}{N_{Re} N_{Sc}^{\frac{1}{3}}} = j_H = \frac{N_{Nu}}{N_{Re} N_{Pr}^{\frac{1}{3}}} = \frac{f}{2}$$

This being described as the Colburn J factor analogy.

It therefore comes down to an understanding of the friction factor characteristics of a system to then have an understanding of the potential mass and heat transfer rates.

This is where the complexity enters the process as there are no universal formula that can be applied to all the applications for which mass and heat transfer can take place.

References to standard texts such as Perry's Chemical Engineers Handbook will advise up to thirty different equations that will relate mass and heat transfer characteristics to that of Reynold's Number.

They are normally separated because the physical layout for process that is designed to allow mass transfer and heat transfer are two different systems.

Shell and tube heat exchanger is very different in design to that of a distillation column.

In systems where you have simultaneous heat and mass transfer such as a dryer or a cooling tower, the heat and mass transfer analogy apply.

In all cases the mass and heat transfer mechanism differs for liquid phase as compared to gas phase and the equations are very different for each.

Gas flowing through a packed column by which the bulk of gas flows across the most of the void passages within the bed is very different to the flow of ribbon of water across the surfaces of random packing.

If we for example use the simplified case of fluid in turbulent flow through a circular pipe

$$j_M = \frac{N_{Sh}}{N_{Re} N_{Sc}^{\frac{1}{3}}} = j_H = \frac{N_{Nu}}{N_{Re} N_{Pr}^{\frac{1}{3}}} = \frac{f}{2} = \frac{0.027}{N_{Re}^{0.2}}$$

We have for mass transfer

$$N_{Sh} = 0.027 N_{Re}^{0.8} N_{Sc}^{0.33} = \frac{k_c \delta x}{D} = 0.027 \left(\frac{dv\rho}{\mu} \right)^{0.8} \left(\frac{\mu}{\rho D} \right)^{0.333}$$

and the equivalent heat transfer

$$N_{Nu} = 0.027 N_{Re}^{0.8} N_{Pr}^{0.33} = \frac{h \delta x}{k} = 0.027 \left(\frac{dv\rho}{\mu} \right)^{0.8} \left(\frac{C_P \mu}{k} \right)^{0.333}$$

These formulae are only true for circumstances where flow is in a smooth pipe and subject to full turbulent conditions. $N_{RE} > 2500$

This correlation in respect to heat transfer is close approximation of the Siedler-Tate equation

$$N_{Nu} = 0.027 N_{Re}^{0.8} N_{Pr}^{0.33} \Phi = \frac{hD}{k} = 0.027 \left(\frac{dv\rho}{\mu} \right)^{0.8} \left(\frac{C_P \mu}{k} \right)^{0.333} \left(\frac{\mu}{\mu_w} \right)^{0.14}$$

Other correlations of interest:

$$N_{Nu} = 1.86 N_{Re}^{0.33} N_{Pr}^{0.33} \kappa \Phi = \frac{hD}{k} = 1.86 \left[\left(\frac{dv\rho}{\mu} \right) \left(\frac{C_P \mu}{k} \right) \left(\frac{d}{L} \right) \right]^{0.33} \left(\frac{\mu}{\mu_w} \right)^{0.14}$$

Siedler-Tate for $50 < N_{RE} < 2100$ again for a smooth pipe

$$N_{Nu} = 0.36 N_{Re}^{0.55} N_{Pr}^{0.33} \Phi = \frac{hD_e}{k} = 0.36 \left(\frac{D_e v \rho}{\mu} \right)^{0.55} \left(\frac{C_P \mu}{k} \right)^{0.33} \left(\frac{\mu}{\mu_w} \right)^{0.14}$$

Kern method for shell side heat transfer coefficient for Shell and Tube heat exchangers.

$$k_c \left(\frac{\rho}{\mu g} \right)^{0.33} = 0.0051 \left(\frac{v\rho}{a_w\mu} \right)^{0.666} \left(\frac{\mu}{\rho D} \right)^{-0.5} (aD_p)^{0.4}$$

$$5,400 < N_{RE} < 1,000,000$$

Mass Transfer liquid side in scrubbing column in packed column with gas counter current to the liquid flow.

$$\frac{k_c}{a_w D} = 5.23 \left(\frac{dv\rho}{\mu} \right)^{0.7} \left(\frac{\mu}{\rho D} \right)^{0.333} (aD_p)^{-2.0} \text{ where } (D_{Packing} > 0.5)$$

$$\frac{k_c}{a_w D} = 2 \left(\frac{dv\rho}{\mu} \right)^{0.7} \left(\frac{\mu}{\rho D} \right)^{0.333} (aD_p)^{-2.0} \text{ where } (D_{Packing} < 0.5)$$

$$5,400 < N_{RE} < 1,000,000$$

Mass Transfer gas side in scrubbing column in packed column with gas counter current to the liquid flow. The coefficient dependent on Packing Size.

9 Contact Surface Area in Real Systems

The effects of mass transfer and heat transfer occurring across a fixed continuous flat surface does not represent the real situation. Mass Transfer occurs across the interface between a gas and liquid phase the liquid phase being distributed across a random packing surface. Heat Transfer occurring across the interface between the inside and outside of a single or set of tubes. Nominating the required surface area isn't practical from the point of view of determining the actual size of the scrubbing tower or the heat exchanger.

Determining the actual physical size of the equipment requires a more practical approach.

In the case of mass transfer, the consideration would be towards the diameter of the column and height of packing. What is required in relation to the packing is knowing what is the generated surface area of the packing per unit volume, leaving the volume of packing as the only variable to be determined.

Standard packings data has been generated for a number of different packing systems to which parameters have been provided.

Packing	a m²/m³
Rashig Ring 1"	190
Pall Ring 1"	206
Berl Saddle 1/2" Ceramic	465
HydraPak 80	93
Pall Ring 2"	108
Tellerettes #1	180
Tellerettes #2	92
Q-Pak Lantec	98.5
Jaeger Tripack 2"	157

This is only a limited sample of the surface area available for engineering bulk random packing. (A detailed list is given in Treybal)

It is noted that when considering random packing types, that the target is to select a specific surface area as high as possible. The smaller the packing the higher the surface area.

The offset is that as the packing size is reduced, the pressure drop increases substantially and the more likely the packing is to blocking up. As with all engineering decisions, it is always a compromise to which packing is selected.

As the condition of random packing scrubber system depend on the interfacial contact of gas and liquid, the rate of mass transfer is dependent on the available surface area of the liquid as it spreads across the surface of the packing. The area of packing that is occupied by liquid is close to that of the total packing surface area, but is not totally. For most design cases a ratio value of 1 is okay for an approximation.

A correlation provided by Onda et al provides good estimation for common packings.

$$\frac{a_w}{a} = 1 - e^{-\delta} \text{ where } \delta = 1.45 \left(\frac{\sigma_c}{\sigma_l} \right)^{0.75} N_{Re}^{0.1} N_{We}^{0.2} N_{Fr}^{-0.05}$$

In terms of the general concept of mass transfer design, knowledge of the specific surface area of the packing is allows the equation to be modified to

$$QdC = -k_c(C - C_0)dA = -k_c a(C - C_0)dV$$

The design therefore comes down to the column packing volume rather than the available surface area.

In the case of Heat Exchangers, the design comes down to simple consideration of the available surface area of the tubing for which heat transfer is taking place with the tube perimeter length x tube length representing the area.

$$Q\rho C_p dT = -h(T - T_0)dA = -hP(T - T_0)dL$$

There a number of crucial points that need to be considered in relation to the straight forward determination of surface area:

- For mass transfer purposes, there is the interface of liquid flowing across the surface of packing in contact with air or gas flowing across a section of packing within a column. This interface surface isn't the same as the actual available packing surface and is always slightly less. $a_w < a$
- The perimeter value for heat transfer is only satisfactory if you only consider simple tube in tube system. Shell and Tube heat exchangers have multiple small tubes within a shell bundle that require more extensive equation to calculate the area per unit length.
- Heat exchangers usually consist of fluids flowing on the outside and inside of a pipe and generally the areas are different. The standard approach is to specify the heat transfer coefficient relative to the outside diameter area and apply the calculation to it.

Heat and mass transfer do not always operate in conditions where someone has specifically designed equipment for it to occur.

Packed and trays columns are always associated with mass transfer, and heat exchangers are always associated with heat transfer.

But a vertical pipe with water flowing across the surface will allow both mass and heat transfer to occur. A rectangular channel of liquid flowing as a result of a gradient will allow mass transfer across the liquid surface. This is the basis for which Hydrogen Sulphide generated in the wall section of the sewer is released into the atmosphere.

10 In conclusion

Molecular diffusion and heat transfer in accordance to Fick's Law and Fourier's law provide a theoretical basis of mass and heat diffusion based on driving forces that operates in nature.

These laws along with newton's law on momentum flow of fluids provide proof of the nature of the transport of matter and energy are closely interrelated and that the mechanism that governs the set conditions of one system govern the conditions of the others.

In real world systems, the effects of turbulence a chaos overcome the molecular diffusion and conduction heat flow that would normally be associated with fluid processes and more sophisticated models involving dimensionless groups need to be considered.

When considering the process conditions under which mass and heat transfer take place, the general condition

$$N_{Sh} = f(N_{Re}N_{Sc}) \text{ and } N_{Nu} = f(N_{Re}N_{Pr})$$

need to be considered for every specific condition and will need to be accounted for in the fact that the formulae will change between laminar and turbulent flows, the liquid phase operating in different conditions to the gas phase conditions. No one general easily applied formula exists and empirical equations based on the design can be obtained from published sources.

11 Variables and Units

Variable	Units	Description
a	m^2/m^3	Specific Surface Areas representing the contact area per unit volume of packing
a_w	m^2/m^3	Wetted Specific Surface Areas
C, C_g	mol/m^3	Molar concentration of diffusing compound liquid / gas phase
C_p	J/kg C	Specific Heat of Fluid
D	m^2/s	Diffusivity of compound in gas or liquid
d, D_e, D_p	m	Pipe Diameter, Equivalent Diameter, Packing Diameter
E	W (watts)	Power gain or loss by way of heat transfer
f	-	Fanning's friction factor
g	m/s^2	Acceleration due to gravity
h	$W/m^2 C$	Heat Transfer Coefficient
j_H, j_M	-	Chilton – Colburn J factor = $f/2$
k_c, k_y, k_x	m/s	Molar mass transfer coefficient. General, gas, liquid phases
k	$W/m C$	Thermal Conductivity of fluid or solid
L	m	Length of pipe
P	m	Perimeter length of pipe
R	m	Radius of pipe
Q	m^3/s	Volumetric Flow
v	m/s	Bulk flow velocity
V	m^3	Volume
T	C	Temperature
ρ	kg/m^3	Bulk Fluid Density
μ, μ_w	kg/ms	Bulk fluid viscosity, wall fluid viscosity
τ, τ_{Wall}	Pa	Shear stress on fluid, wall shear stress

x, x_i	mol/mol	Molar concentration liquid phase bulk & interface values
y, y_i	mol/mol	Molar concentration gas phase bulk & interface values
$\Delta X, \Delta Y, \Delta Z$	m	Cartesian co-ordinate incremental distance values.
σ_L, σ_C	N/m	Liquid Surface Tension and Critical Surface Tension

12 Literature Cited in this Document

- Perry's Chemical Engineer's Handbook 7th Ed – R H Perry and W D Green
- Principles of unit operations 2nd Ed - A. Foust et al
- Mass Transfer Operations 4th Ed – R E Treybal
- Process Heat Transfer – Q D Kern
- Mass transfer coefficients between gas and liquid phases in packed columns. Journal of Chemical Engineering – Japan, Onda, K. H.; Takeuchi, H.; Okamoto, Y.