

AIR POLLUTION CONTROL, WASTE WATER AND SOLID WASTES MANAGEMENT BMEGEÁTAG04

4th part lecture handout
AIR POLLUTION CONTROL: "Particle removal from gases"

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subject's website
http://www.ara.bme.hu/oktatas/tantargy/NEPTUN/BMEGEATAG04/ENGLISH_course

Contents, main topics of the lecture:

1. Characterisation of particle-laden mixtures. Aerosols.
2. Particle dynamics: equation of motion for particle-laden flows and for particle
3. Measurement of particle concentration
4. Gas cleaning equipments, various types of particle separators

1. Characterisation of particle-laden mixtures. Aerosols

Aerosols

Definition of aerosols:

Aerosols are defined as being gas & particle mixtures in quasi-stable state. The mixture contains partly gas as primary (or carrier) phase and partly solid / liquid particulate matter (as secondary phase).

The "quasi-stable state" means that the characteristics (e.g. particle number, mass concentration) of the mixture in a given volume do not change significantly i.e. "nearly stable" in time.

Change of the characteristics of the mixture can occur due to the:

- settling out of larger particles from the given volume of the mixture, or
- diffusion and agglomeration of the small particles.

Both may cause increase or decrease of total mass of the particles in the fixed volume, hence may cause changing of the characteristics of the mixture.

Diameter (x [μm]) range of the particles in aerosols: $0,01 \mu\text{m} \leq x \leq 50 \mu\text{m}$

Note, that the lower & upper limiting values are not strict limiting values: "0,01 μm " & "50 μm " means that approx. a few hundreds & few times ten microns can be considered as the lower & upper limits of the diameter range of aerosols.

Note: $1\mu\text{m} = 10^{-3} \text{ mm} = 10^{-6} \text{ m}$

Examples: The resolution of sensitivity of a human finger tip is about 40 microns.

The human hair's diameter is between 40÷100 microns.

The average height of surface roughness of a bearing ball is approx. 0,01 micron.

The diameter of seeding particles for Laser Doppler Velocimetry and flow visualization (oil smoke or spherical oil droplets from fog generator) is approx. 1÷3 microns.

Types and sizes of particles

Dust: $x \geq 0,2 [\mu\text{m}]$

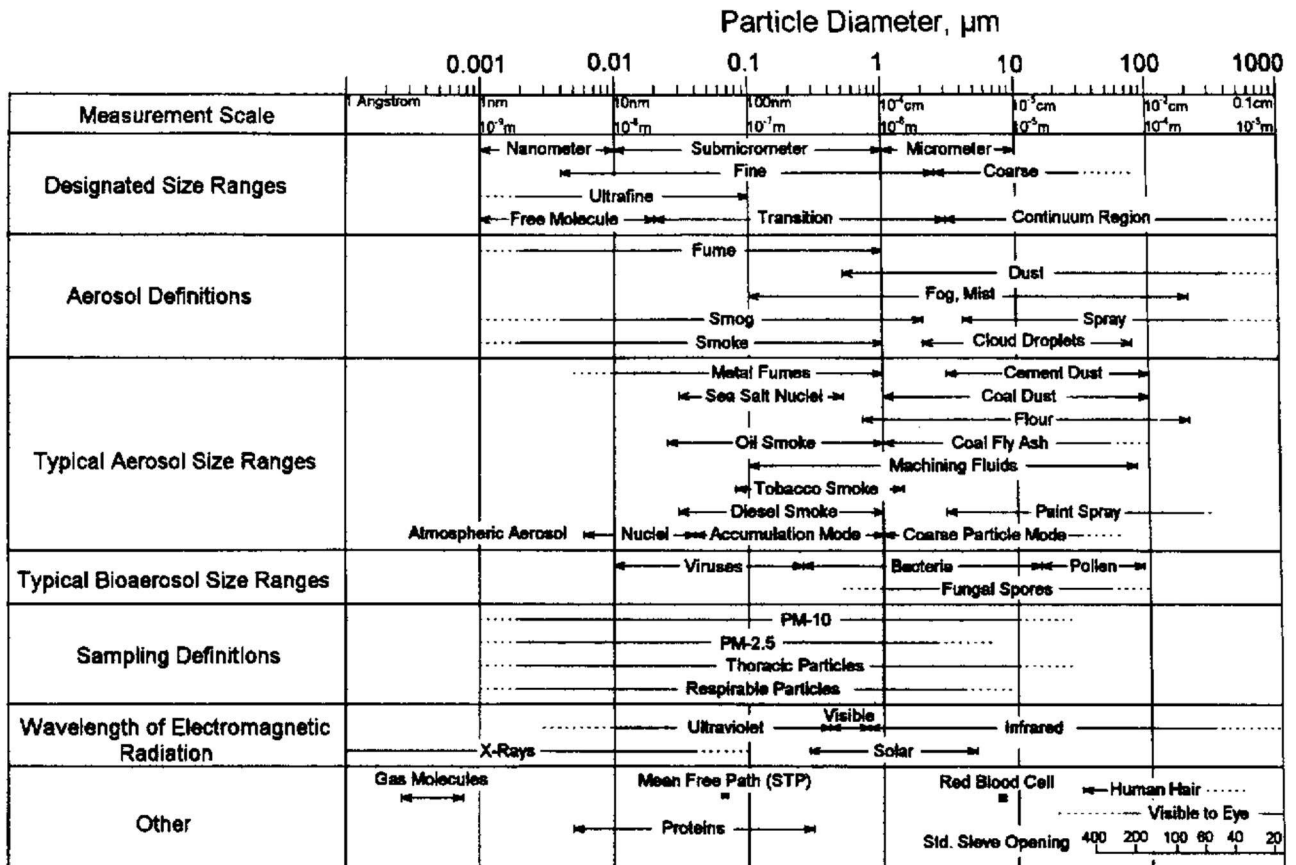
description: **solid** particles, produced by breaking or attrition, abrasion, wearing of solid substances, perceptible to the eye, the diameter is larger than the wave length of light.

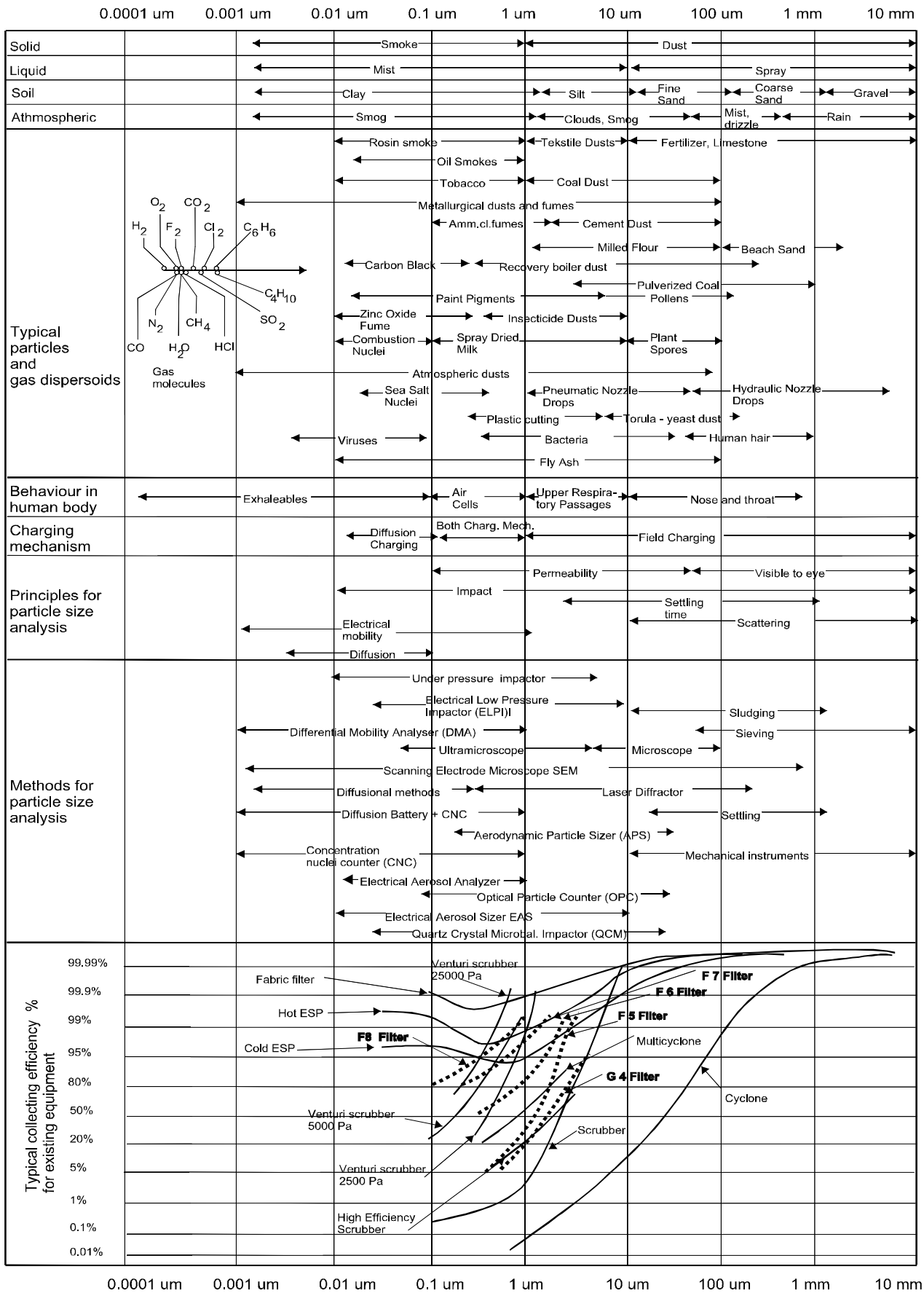
Smoke /fume: $x \leq 1 [\mu\text{m}]$

description: **solid or liquid** particles or droplets, originated from condensation or chemical reaction, in most cases chain-like structures. Produced at combustion, chemical processes etc.

Mist /fog: $0,1 \leq x \leq 200 [\mu\text{m}]$

description: **liquid** droplets originated from steam condensation or by atomization, spraying. Droplets in the mist are in equilibrium state with their saturated steam.





Size of particles:

NOTE: In case of spherical particles the diameter is denoted here by x .

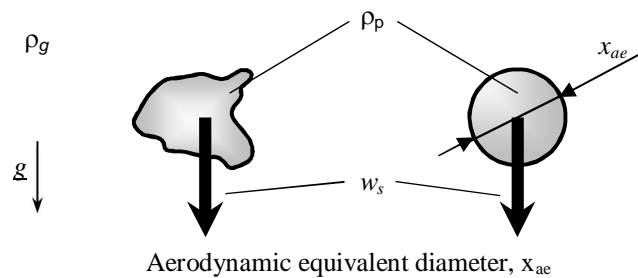
Elsewhere, and sometimes also here in the handout the *diameter* of a *particle* is usually denoted by d_p .

How to characterize - define - the “size” of non-spherical particles? By practical reason it is needed to introduce the so-called equivalent diameter that is the diameter of an ideal spherical particle from the same material as the real particle of arbitrary shape.

Various types of equivalent diameter can be defined based on:

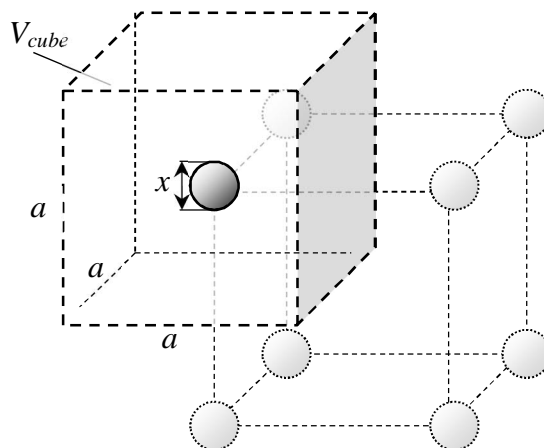
- geometrical equivalence (based on the particle surface, volume or projected area) / used e.g. in chemical reaction calculation/
- **aerodynamic equivalence (see below in more detailed)**
- optical equivalence (based on refractive index) /in optical concentration measurement devices/
- electrical equivalence (based on electric conductivity, resistance) /used when concentration measurement is done by electric resistivity meas. devices/.

In particle dynamics the most relevant is to deal with the x_{ae} **aerodynamic equivalent diameter**, that is defined to be the diameter of a spherical particle from the same material (ρ_p =same) as the real particle, settling with the same w_s settling velocity in the same gas (ρ_g =same).



Average relative distance (a/x) between neighboring particles in gas:

Let's calculate the c [kg/m^3] mass concentration of n particles evenly distributed in a particle-gas mixture having a volume of V_{g+p} . Let's assume that each particle is sitting in the center of a cube. (see Figure below).



The concentration can be calculated:

$$c = \frac{\sum m_p}{V_{g+p}} \cong \frac{\sum m_p}{V_{g+p}} = \frac{\sum V_p \cdot \rho_p}{V_{g+p}} = \frac{n \cdot \frac{x^3 \cdot \pi}{6} \cdot \rho_p}{n \cdot a^3} = \frac{\frac{x^3 \cdot \pi}{6} \cdot \rho_p}{a^3}$$

where c [kg/m^3] mass concentration, a [m] average distance between particles, ρ_p [kg/m^3] density of particles, n is number of particles.

For the average relative distance (a/x) between neighboring particles in gas we get:

$$\frac{a}{x} = \sqrt[3]{\frac{\rho_p \cdot \pi}{6 \cdot c}}$$

Example:

If we consider monodisperse particle distribution where all particles have $x=3 \mu\text{m}$ diameter with $\rho_p=2000\text{kg/m}^3$, and $c = 10 \text{ g/m}^3$ (that is relatively very high concentration of particles), the $a/x = 47$.

a) It means that aerosols are very **dilute mixtures**: the neighboring particles are far from each other (approx. $a=5\text{m}$ when $x=10\text{cm}$ would be the particle diameter). Hence, the possibility of collision, momentum exchange between two particles is relatively small.

b) In 1cm^3 volume there are 350000 particles for $c=10 \text{ g/m}^3$. Even in case of small $0,1 \text{ g/m}^3$ concentration we get 3500 particle in 1cm^3 volume. Notwithstanding that it is a **dilute mixture** the number of particles is very high in given volume even for low concentration. That is important to know when very strict demand is defined on air quality (e.g. at surgery rooms, when dealing with toxic or infective particles is concerned)

$c[\text{g/m}^3]$	$\frac{a}{x}$	N [db/cm ³]
10	47	350.000
1	101	35.000
0.1	218	3.500

Conclusion:

- in case of usual particle concentration values the particle-laden flows are very dilute mixtures. (the distance between neighboring particles is very large).
- particles are present with very high number even in particle-gas mixtures having very low concentration.

Characterization of particle assembly:

Particle size distribution curves:

Considering polydisperse particle distribution with size range of $x_{\min} < x < x_{\max}$

Cumulative or undersize distribution related to number of particles: $Q_0=N/N_{\text{tot}}=f(x)$. Q_1, Q_2, Q_3 .

Subscript denotes: 0: related to number of...

1: related to length of... (1D – one dimensional quantity)

2: related to surface of...(2D – two dimensional quantity)

3: related to volume or mass of...(3D – three dimensional quantity)

If $Q_0=N/N_{\text{tot}}=f(x)$ and the overall number of particles N_{tot} are known, the number of particles in the size range between $[x]$ and $[x+\Delta x]$ can be calculated:

$$\Delta N = N_{\text{tot}} \cdot \frac{dQ_0}{dx} \cdot \Delta x$$

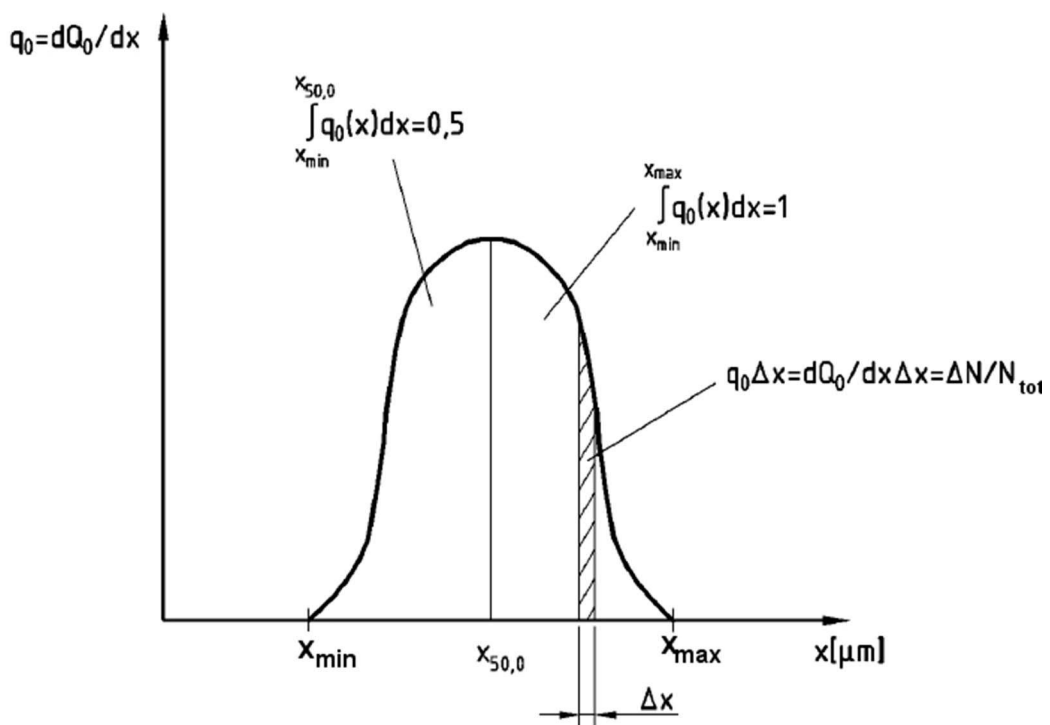
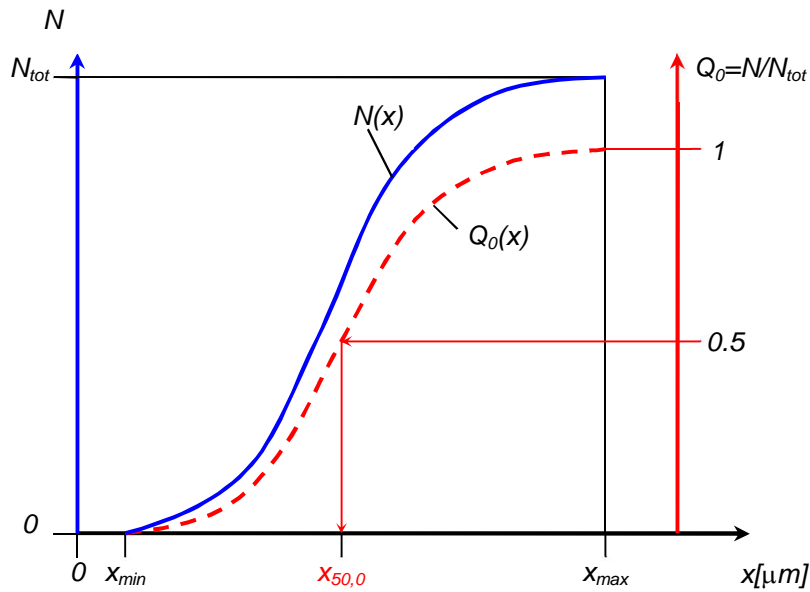
Cumulative or undersize distribution related to number of particles: Q_0 . The value of $Q_0(x)$ for given x gives information how many percentage of N_{tot} particles have smaller diameter than x .

Taking the tangent of the Q_0 curve (see Figure below) is denoted by q_0 (density function):

$$q_0 = \frac{dQ_0}{dx}$$

we get for the number of particles in the range between $[x]$ and $[x + \Delta x]$:

$$\Delta N = N_{tot} \cdot \frac{dQ_0}{dx} \cdot \Delta x = N_{tot} \cdot q_0 \cdot \Delta x$$

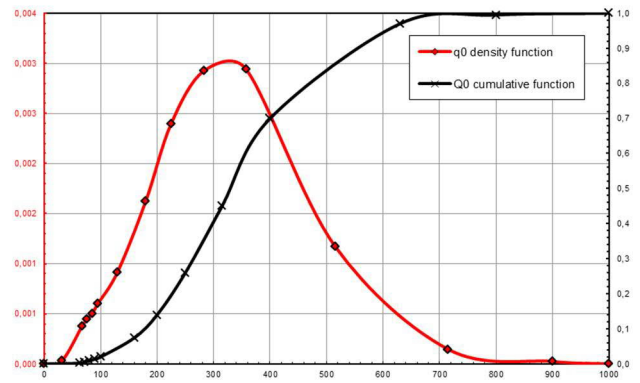
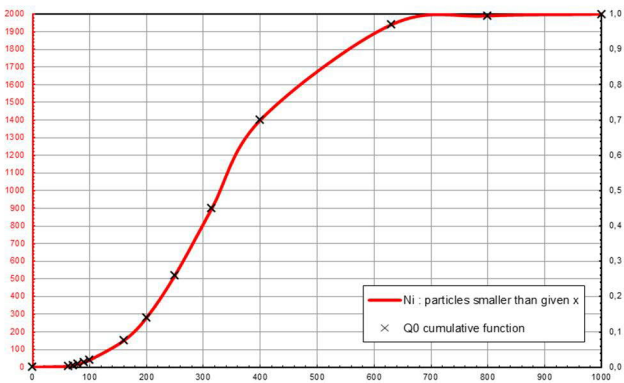


When q_0 cumulative distribution related to the particle number is given (see Figure below), we can obtain the average diameter of the particle distribution (\bar{x}_0) related to particle number:

$$\bar{x}_0 = \frac{1}{N_{tot}} \int_{x_{min}}^{x_{max}} x \cdot N_{tot} \cdot q_0 dx = \int_{x_{min}}^{x_{max}} x \cdot q_0 dx$$

EXAMPLE for a $N_{tot}=2000$ particle ensemble.

$N_{tot}=2000$ particles					$Q_0=$	$q_0=$	$x_{0,mean}=$		
x_i μm	Δx_i μm	$x_{i,mean}$ μm	N_i particles	ΔN_i particles	N_i/N_{tot} -	$\Delta Q_0/\Delta x_i$ -	$\sum x_{i,mean} q_{0i} \Delta x_i$ $x_{1,0}$	$x_{2,0}$	$x_{3,0}$
0	0		0		0,00	0			
63	63	31,5	4	4	0,00	0,0000	0,1	2,0	62,5
71	8	67	10	6	0,01	0,0004	0,2	13,5	902,3
80	9	75,5	18	8	0,01	0,0004	0,3	22,8	1721,5
90	10	85	28	10	0,01	0,0005	0,4	36,1	3070,6
100	10	95	40	12	0,02	0,0006	0,6	54,2	5144,3
160	60	130	150	110	0,08	0,0009	7,2	929,5	120835,0
200	40	180	280	130	0,14	0,0016	11,7	2106,0	379080,0
250	50	225	520	240	0,26	0,0024	27,0	6075,0	1366875,0
315	65	282,5	900	380	0,45	0,0029	53,7	15163,2	4283600,5
400	85	357,5	1400	500	0,70	0,0029	89,4	31951,6	11422683,6
630	230	515	1940	540	0,97	0,0012	139,1	71610,8	36879536,3
800	170	715	1990	50	1,00	0,0001	17,9	12780,6	9138146,9
1000	200	900	2000	10	1,00	0,0000	4,5	4050,0	3645000,0
						1,000	352	381	407
							$x_{1,0}$	$x_{2,0}$	$x_{3,0}$
							[μm]	[μm]	[μm]



ps.: There is a simple relation/conversion between distributions related to various quantities e.g. conversion of Q_0 to Q_3 :

$$Q_3(x) = \frac{\int_{x_{min}}^x x^3 \frac{\pi}{6} N_{tot} \frac{dQ_0}{dx} dx}{\int_{x_{min}}^{x_{max}} x^3 \frac{\pi}{6} N_{tot} \frac{dQ_0}{dx} dx} = \frac{\int_{x_{min}}^x x^3 q_0 dx}{\int_{x_{min}}^{x_{max}} x^3 q_0 dx}$$

2. Particle dynamics

Effect of particles on the gas flow field

The Navier-Stokes equation of motion for the carrier gas flow (primary phase) is extended by the influence of the particles (secondary phase). The effect of particles on the gas flow is given by the force (\underline{t}).

$$\frac{\partial \underline{v}}{\partial t} + \text{grad} \frac{v^2}{2} - \underline{v} \times \text{rot} \underline{v} = \underline{g} - \frac{1}{\rho_g} \text{grad} p + \nu \Delta \underline{v} + \underline{t}$$

where \underline{v} [m/s] is gas velocity vector, p [Pa] is pressure, \underline{g} [N/kg] is gravity field strength vector, ν [m²/s] is kinematic viscosity and.

The last term on the right hand side is \underline{t} [N/kg_{gas}], the force acting on the gas from particles in 1 kg of gas:

$$\underline{t} = - \frac{n \underline{F}}{\rho_g}$$

n [particles/m³] : particle number concentration

\underline{F} [N / particle] : aerodynamic force acting on one particle

ρ_g [kg/m³] : gas density

This extended form of Navier-Stokes equation is very difficult to solve in case of aerosols, since \underline{t} is to be determined for a polydispersed real-shaped particulate phase where the force for each particle is needed to determine first, that is impossible.

Let's examine when would it be possible to neglect this term \underline{t} ?

Assuming sphericity for the particles, when taking the ratio of the total volume of the particles and the total volume of the gas in a given control volume of the aerosol, we will get the dimensionless α [-] volume ratio of the phases.

Volume ratio:

$$\alpha_p = \frac{V_p}{V_g} = \frac{\frac{d_p^3 \pi}{6}}{a^3} = \frac{c_p}{\rho_p}$$

where c_p [kg/m³] is the mass concentration of the particles, ρ_p [kg/m³] is the density of the particles.

Taking the ratio of the total mass of the particles and the total mass of the gas in a given control volume of the aerosol, we get the dimensionless M [-] mass loading ratio.

Mass loading ratio:

$$M = \frac{c_p}{\rho_g} = \alpha_p \frac{\rho_p}{\rho_g}$$

$$\frac{M}{\alpha_p} = \frac{\rho_p}{\rho_g}, \text{ or } \frac{\alpha_p}{M} = \frac{\rho_g}{\rho_p}$$

where:

c_p : particle mass concentration

ρ_g : density of gas (carrier phase)

ρ_p : density of particle (material)

For example: typical values of volume ratio and mass loading ratio can be seen in tables below.

α_p		$\rho_p [kg/m^3]$		
		800	1500	2500
$C_p [g/m^3]$	0,0001	$1,3 \cdot 10^{-10}$	$6,7 \cdot 10^{-11}$	$4,0 \cdot 10^{-11}$
	0,001	$1,3 \cdot 10^{-9}$	$6,7 \cdot 10^{-10}$	$4,0 \cdot 10^{-10}$
	0,01	$1,3 \cdot 10^{-8}$	$6,7 \cdot 10^{-9}$	$4,0 \cdot 10^{-9}$
	0,1	$1,3 \cdot 10^{-7}$	$6,7 \cdot 10^{-8}$	$4,0 \cdot 10^{-8}$
	1	$1,3 \cdot 10^{-6}$	$6,7 \cdot 10^{-7}$	$4,0 \cdot 10^{-7}$
	10	$1,3 \cdot 10^{-5}$	$6,7 \cdot 10^{-6}$	$4,0 \cdot 10^{-6}$
	100	$1,3 \cdot 10^{-4}$	$6,7 \cdot 10^{-5}$	$4,0 \cdot 10^{-5}$

M		$\rho_g [kg/m^3]$		
		0,8	1,0	1,2
$C_p [g/m^3]$	0,0001	$1,3 \cdot 10^{-7}$	$1,0 \cdot 10^{-7}$	$8,3 \cdot 10^{-8}$
	0,001	$1,3 \cdot 10^{-6}$	$1,0 \cdot 10^{-6}$	$8,3 \cdot 10^{-7}$
	0,01	$1,3 \cdot 10^{-5}$	$1,0 \cdot 10^{-5}$	$8,3 \cdot 10^{-6}$
	0,1	$1,3 \cdot 10^{-4}$	$1,0 \cdot 10^{-4}$	$8,3 \cdot 10^{-5}$
	1	$1,3 \cdot 10^{-3}$	$1,0 \cdot 10^{-3}$	$8,3 \cdot 10^{-4}$
	10	$1,3 \cdot 10^{-2}$	$1,0 \cdot 10^{-2}$	$8,3 \cdot 10^{-3}$
	100	$1,3 \cdot 10^{-1}$	$1,0 \cdot 10^{-1}$	$8,3 \cdot 10^{-2}$

The M mass loading ratio for typical concentration for dusty gases (upstream of particle separators) has a value of at least 10^{-1} . Notwithstanding, that the number concentration is high, the mass loading ratio is very small. The particle's forces \underline{t} , hence the effect of particle phase on the gas flow field can be neglected when

- the mass loading ratio $M \ll 1$, i.e. $\frac{c}{\rho_g} \ll 1$,

and

- the particle acceleration is in the same order of magnitude as the carrier gas-phase acceleration $\frac{d\underline{u}_p}{dt} \approx \frac{d\underline{v}}{dt}$.

Therefore, in cases when $\frac{c}{\rho_g} \frac{d\underline{u}_p}{dt} \ll \frac{d\underline{v}}{dt}$, the effect of particle phase on the flow field can be

neglected. From other viewpoint, we can state that one single particle moving in the gas in this dilute mixture (aerosol) cannot change significantly the gas' momentum, the effect of particles can be neglected.

It is worth to mention here that the particle's pathline due to the drag is always influenced by the carrier gas flow, see in next chapter: defining the aerodynamic (drag) force acting on the particles.

Aerodynamic (drag) force acting on a single particle:

The particle Reynolds number ($Re_p = \frac{w \cdot x}{\nu}$) is small, viscosity is dominant in a fluid flow around the particle (using the relative co-ordinate system fixed to the particle), see calculated results for flow field around a sphere for various particle Reynolds-number (Source: MICHAELIDES: Particles, Bubbles, & Drops (2006) p.109.)

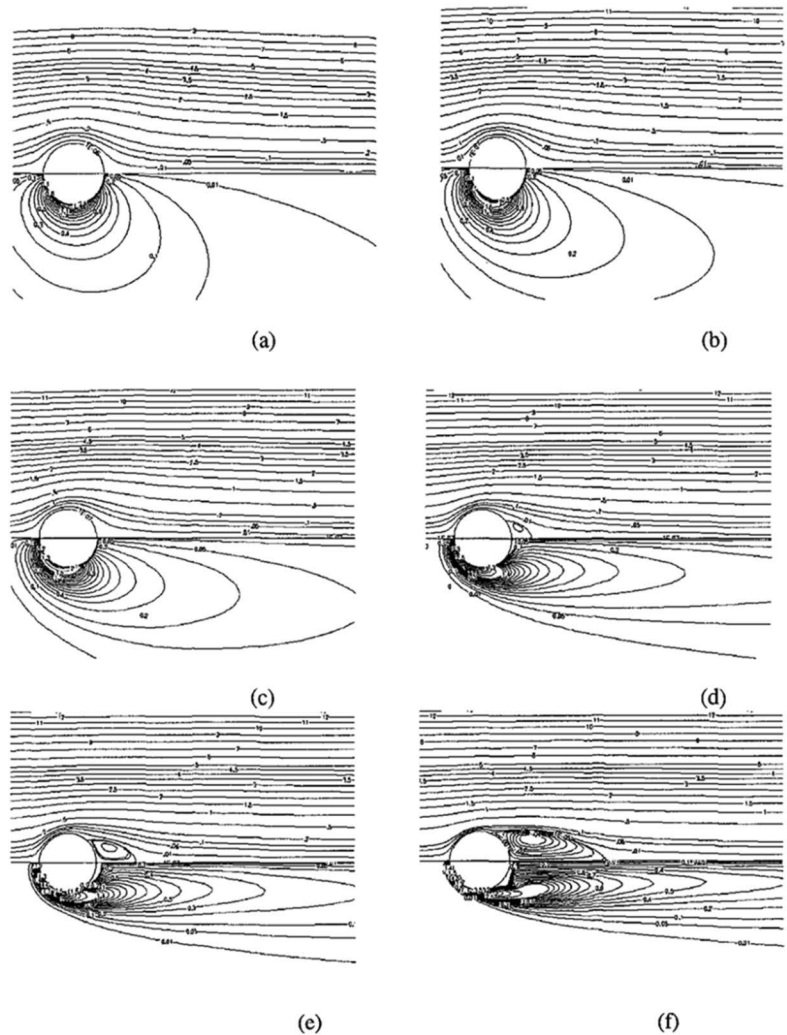


Fig. 4.1 streamlines and vorticity around a rigid sphere. The six parts correspond to $Re=1, 5, 10, 50, 100$ and 500 . The top half depicts the streamlines and the bottom half the vorticity contours

Stokes assumed the following: if $Re_p < 0.1$, the well known Stokes relation for F_d drag force acting on a particle (having particle diameter x) moving with w relative velocity in the gas (μ is dynamic viscosity) is:

$$\underline{F}_d = \underline{F}_{Stokes} = 3 \cdot \pi \cdot \mu \cdot x \cdot \underline{w}$$

where $\underline{v} = \underline{u} + \underline{w}$

\underline{v} absolute (gas) velocity

\underline{u} particle velocity

\underline{w} relative velocity

The c_d drag coefficient of the particle is defined by the relative force to the dynamic pressure of the relative flow and the reference cross-section of the spherical particle:

$$c_d = \frac{|\underline{F}_d|}{\frac{\rho}{2} w^2 \cdot A_{\text{ref}}} = \frac{|\underline{F}_d|}{\frac{\rho}{2} w^2 \frac{\pi}{4}}$$

Substituting the Stokes relation for \underline{F}_d into equation of c_d , we get a very simple form for the drag coefficient of a sphere using particle Reynolds-number:

$$c_d = \frac{24}{\text{Re}_p}$$

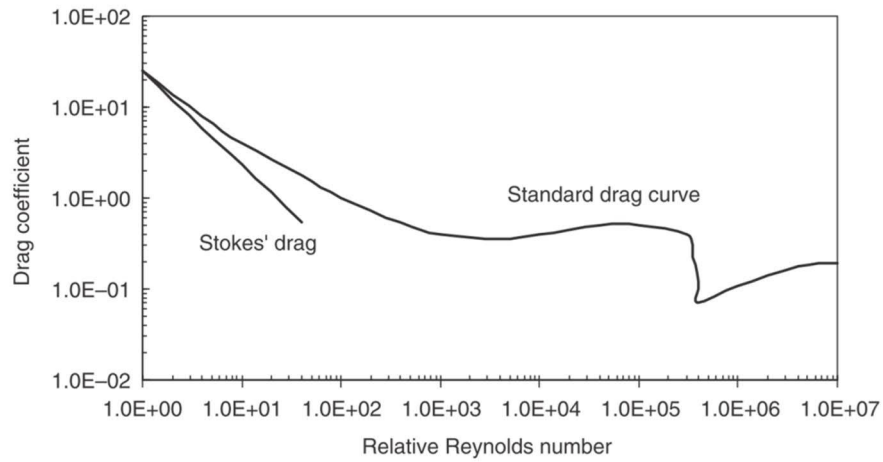
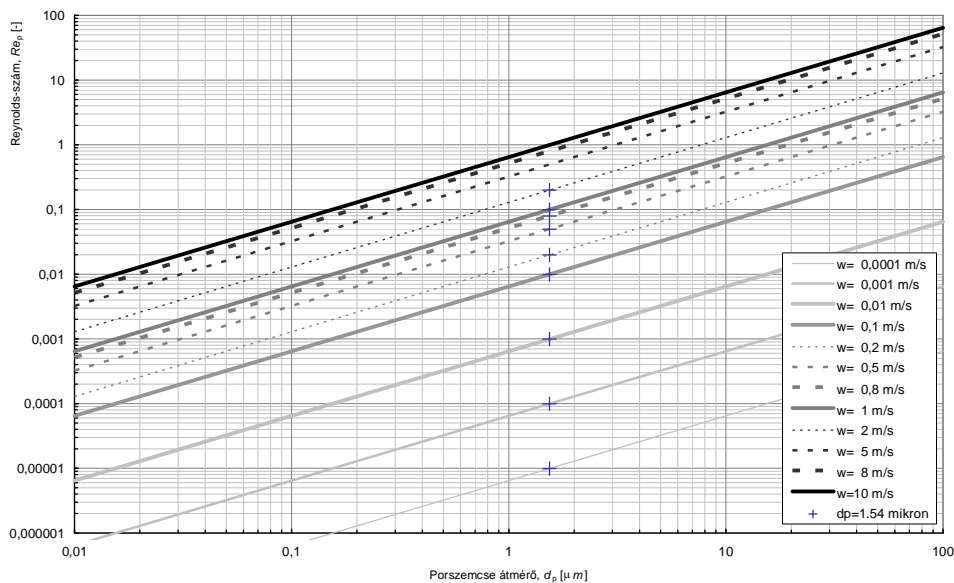


FIGURE 1.9 The standard drag coefficient curve for a solid sphere in steady flow.

Typical particle Reynolds number is plotted (see diagram below) for particles having diameter between $0.01 \mu\text{m} < d_p < 100 \mu\text{m}$, and for relative velocity $10^{-4} \text{m/s} < w < 10 \text{m/s}$. That ensures that Stokes' drag form can be used in the typical engineering applications when particles are separated from aerosols.



The Stokes' drag form can be used only in cases when $\text{Re}_p < 0.1$ (or better for cases when $\text{Re}_p < 0.01$). Researchers in this field obtained various corrected forms for Stokes' drag coefficient for higher

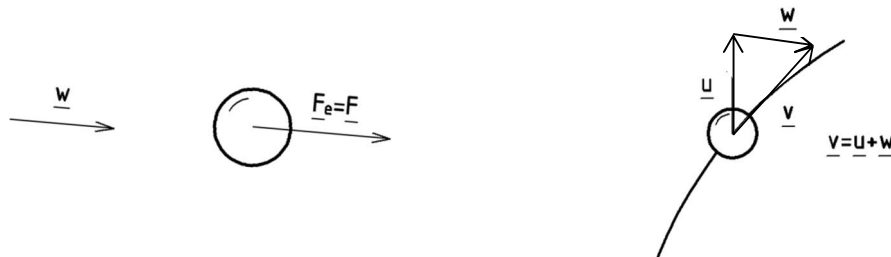
particle Reynolds numbers, since the real drag differs from this Stokes' drag for higher relative velocities or larger particles. Based on experiments and/or CFD numerical simulation studies the following expressions can be used for drag coefficient, e.g.

a) Oseen's drag:
$$c_d = \frac{24}{Re_p} \left(1 + \frac{3}{16} Re_p \right) \quad \text{for } Re_p < 5.$$

b) Michaelides's drag:
$$c_d = \frac{24}{Re_p} \cdot (1 + 0,15 \cdot Re_p^{0,687}) \quad \text{for } 0.1 < Re_p < 1000.$$

Momentum equation for particles moving with \underline{w} relative velocity in gas flow

The particle drag force vector (\underline{F}_d , denoted in figure below by $\underline{F}_e = \underline{F}$) is defined by the direction of the relative velocity \underline{w} . The particle's relative velocity to the gas flow is $\underline{w} = \underline{v} - \underline{u}$.



Due to Newton's 2nd law the particle's momentum equals to the sum of the acting forces. Forces: gravity force and drag force. (Usually we may neglect the bouyancy force when solid/liquid partilces are dispersed in gaseous fluid flow).

$$m_p \frac{d\underline{u}}{dt} = \underline{F}_g + \underline{F}_d$$

Assuming a spherical particle with a diameter x , and substituting Stokes' drag, we get:

$$\frac{x^3 \pi}{6} \rho_p \frac{d\underline{u}}{dt} = \frac{x^3 \pi}{6} \rho_p \underline{g} + 3\pi \mu x \underline{w}$$

Dimensionless equation of motion of the particle:

As a usual non-dimensionalizing formulation procedure let's multiply the equation with $\frac{l_0}{v_0^2}$, where

l_0 is a characteristic length (e.g. $l_0 = x$), and v_0 is a characteristic velocity (e.g. $v_0 =$ average gas flow velocity)

$$\frac{x^3 \pi}{6} \rho_p \frac{d\underline{u}}{dt} = \frac{x^3 \pi}{6} \rho_p \underline{g} + 3\pi \mu x \underline{w} \quad \left| \cdot \frac{l_0}{v_0^2} \right.$$

Then we get the following dimensionless form of equation of motion:

$$\frac{d \frac{\underline{u}}{v_0}}{d \frac{t}{l_0/v_0}} = \frac{\underline{g} l_0}{v_0^2} + \frac{18 \mu}{x^2 \rho_p} \frac{l_0}{v_0} \frac{\underline{w}}{v_0} .$$

Dimensionless momentum equation for particles (' denotes dimensionless quantities, e.g. $u' = \frac{u}{v_0}$).

$$\frac{du'}{dt'} = \frac{\underline{g} l_0}{v_0^2} + \frac{18 \mu}{x^2 \rho_p} \frac{l_0}{v_0} \frac{w'}{v_0}$$

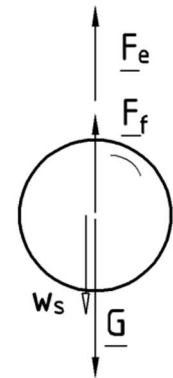
Before rewriting this form of dimensionless momentum equation, the settling velocity of the particle is needed to define.

Settling velocity of the particle (w_s):

Settling of particle of ρ_p density in a still ambient of gas having ρ_g density calls for an equation of motion for the particle settling with a constant settling velocity of w_s . \underline{G} is the gravity force; drag force is \underline{F}_d - denoted in figure below again by \underline{F}_e . Here we do not neglect the \underline{F}_f buoyancy force:

$$\frac{x^3 \pi}{6} \rho_p g = \frac{x^3 \pi}{6} \rho_g g + 3\pi \mu x w_s$$

Settling velocity: $w_s = \frac{x^2 (\rho_p - \rho) g}{18\mu}$ if $\rho_p \gg \rho \Rightarrow w_s = \frac{x^2 \rho_p g}{18\mu}$.

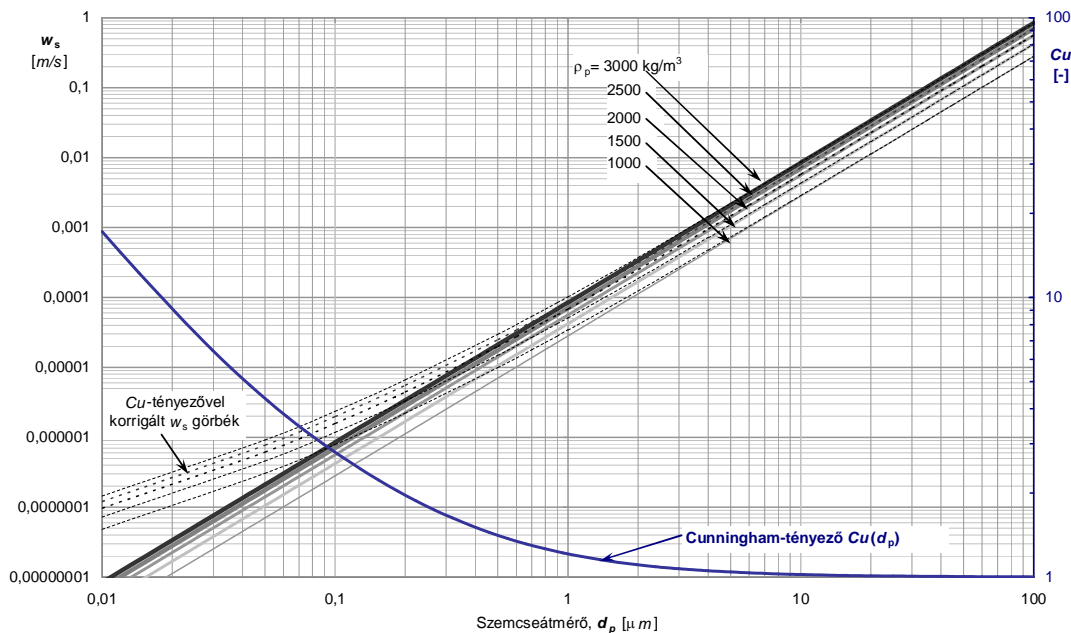


Correction of settling velocity due to the diffusion effect in submicron particle size-range is done by the Cunningham–correction factor (Cu):

$$w_{s,corr} = Cu \cdot w_s$$

where $Cu = 1 + \frac{2 A \lambda}{x}$ is the Cunningham coefficient (or Cunningham correction factor),

where $A \approx 1.4$, and λ is the mean free path of molecules, at room-temperature $\lambda = 6.5 \times 10^{-2} \mu m$.



w_s settling velocity (cont. lines) ad $w_{s,corr}$ corrected settling velocity (dashed lines)
Settling velocity as function of particle diameter and density

Using the settling velocity now we can rewrite the dimensionless momentum equation for particles in the following form.

Firstly, by neglecting the effect of the gravity field strength the dimensionless equation of motion of particles will turn to another form using the w_s settling velocity:

$$\frac{du'}{dt'} = \frac{l_0}{v_0^2} g + \frac{18 \mu l_0}{x^2 \rho_p v_0} w' \cong \frac{18 \mu l_0}{x^2 \rho_p v_0} w' = \frac{g \cdot l_0}{w_s v_0} w'$$

$$\frac{du'}{dt'} = \frac{g \cdot l_0}{w_s v_0} \underline{w}'$$

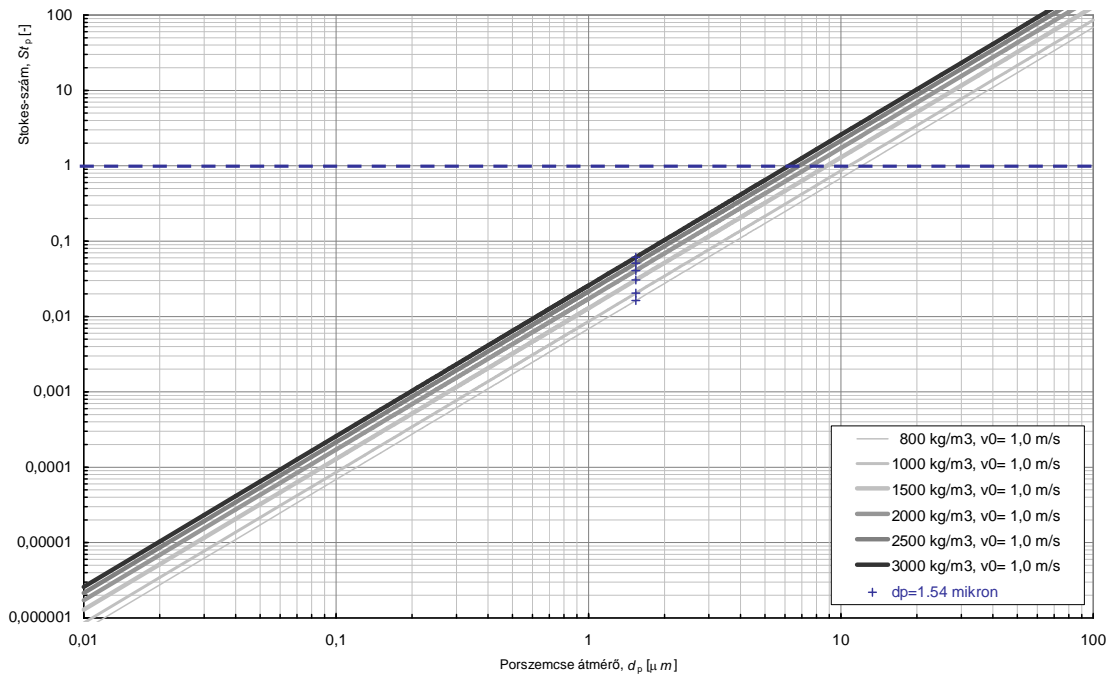
Introducing Ψ_p inertia parameter will help us to evaluate the particle motion in gas flow:

$$\Psi_p = \frac{w_s v_0}{g \cdot l_0}$$

Inertia parameter of particles Ψ_p is also called as the particle Stokes-number, St_p . The particle Stokes number is usually defined in CFD by the relative time scales of the particle and gas flow field.

$$\frac{w_s v_0}{g \cdot l_0} = \Psi_p \equiv St_p = \frac{\tau_p}{\tau_g}$$

Let's calculate inertia parameter e.g. for particles ($0.01 \mu\text{m} < d_p < 100 \mu\text{m}$) in gas having density between $800 \text{kg/m}^3 < \rho_p < 3000 \text{kg/m}^3$ while using $v_0 = 1 \text{m/s}$.

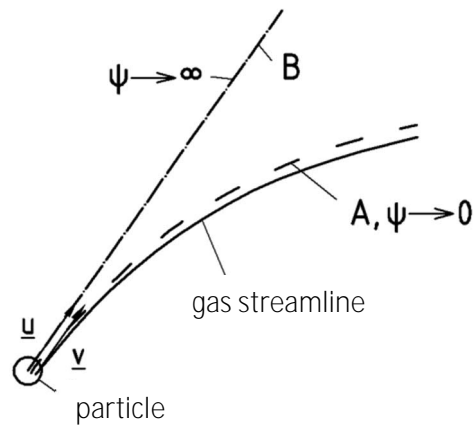


Inertia parameter (Ψ_p) /or particle Stokes number (St_p) vs. particle diameter

Introducing inertia parameter into the dimensionless momentum equation for particles we get:

$$\frac{du'}{dt'} = \frac{1}{\Psi_p} \underline{w}' = \frac{1}{\Psi_p} (\underline{v}' - \underline{u}')$$

Now we can examine how particle move in gas flow depending on their inertia parameter.

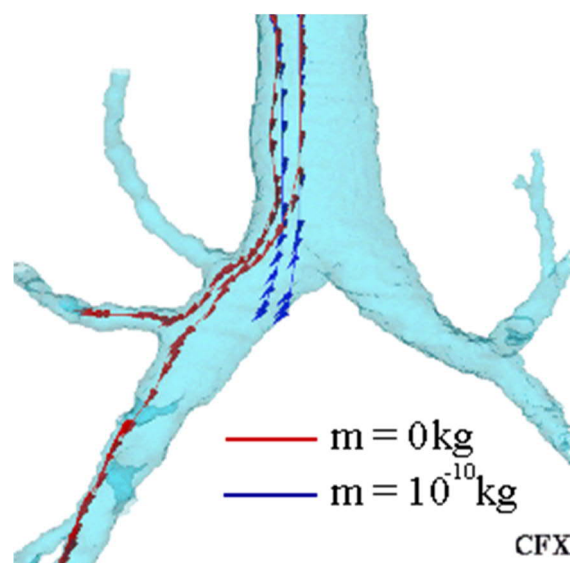


case A) dashed line in the upper figure: When $\psi \rightarrow 0$, for small (x is small) and/or light (ρ_p is small) particles, which settling velocity is small, or $w_s \rightarrow 0$, and if $(\underline{v}' - \underline{u}') \neq 0 \Rightarrow \frac{d\underline{u}'}{dt'} \rightarrow \infty$, hence particle move along the gas streamline, particle follow the carrier gas flow.

case B) dash-dot line in the upper figure: When $\psi \rightarrow \infty$, for large and/or heavy particles, which settling velocity is large, $\frac{1}{\psi} \rightarrow 0$, consequently $\frac{d\underline{u}'}{dt'} \rightarrow 0$. hence particle move along its initial path, leaving the gas streamline.

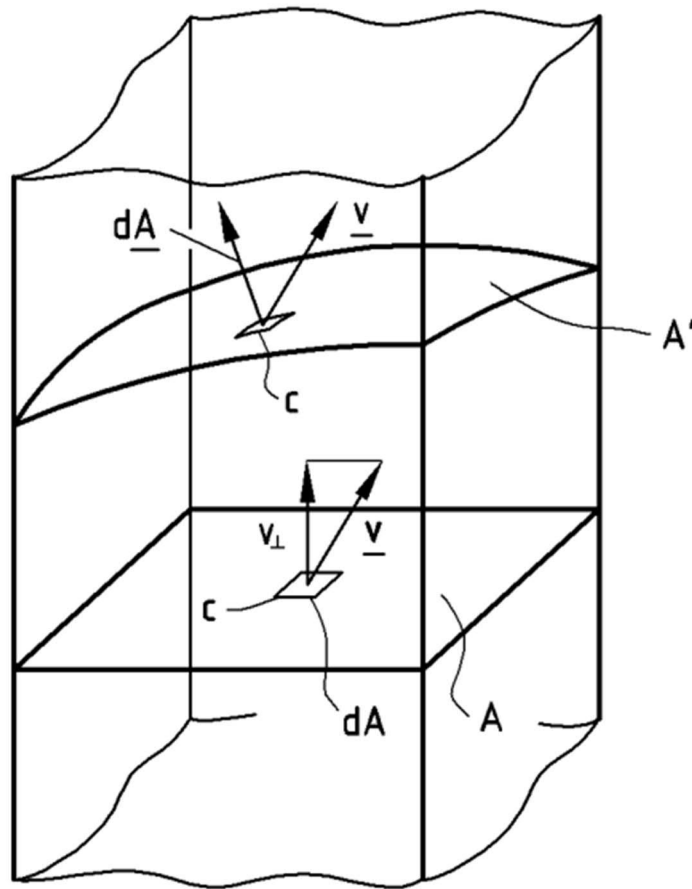
Numerical simulation (CFD) result of particle pathlines in human bronchial duct-segment are shown in figure below:

- RED LINE: „weightless” ($m=0\text{kg}$, $\Psi=0$) particle, this particle fully follows the inhaled air streamline, even in the high curved Y-ducts,
- BLUE LINE: a particle having a diameter of approx. $d_p=40\mu\text{m}$ ($m=10^{-10}\text{kg}$, $\Psi \approx 250$) leaves the inhaled air streamline in the first Y junction.



3. Measurement of concentration

Measurement of mean dust concentration and dust mass flow rate in a duct.



Mean dust concentration \bar{c} [kg/m³]

$$\bar{c} = \frac{\int_{A'} c \underline{v} \, dA}{\int_{A'} \underline{v} \, dA}, \quad \bar{c} = \frac{\int_A c v_{\perp} \, dA}{\int_A v_{\perp} \, dA}, \quad \bar{c} \approx \frac{\sum_{i=1}^n c_i v_{\perp i} \Delta A_i}{\sum_{i=1}^n v_{\perp i} \Delta A_i}$$

where v_{\perp} is the velocity component perpendicular to the A cross section.

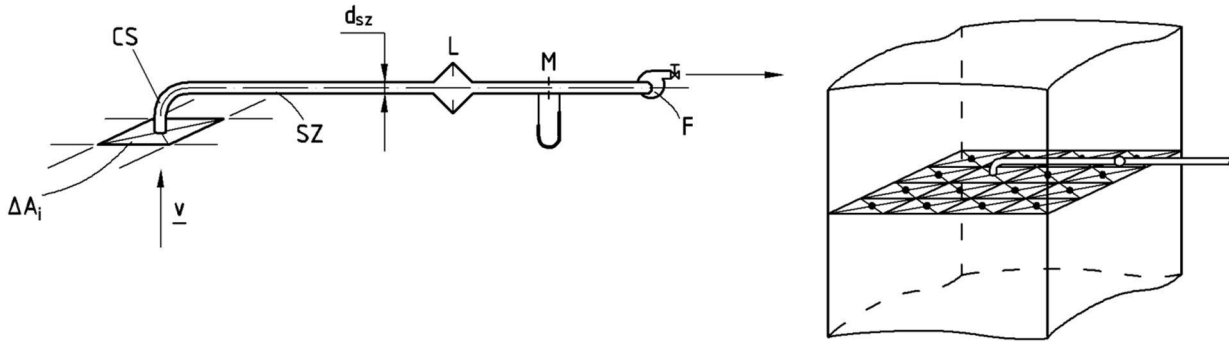
If each $\Delta A = A/n$, i.e. the cross section is divided into n measuring cross section:

$$\bar{c} \approx \frac{\sum_{i=1}^n c_i v_{\perp i}}{\sum_{i=1}^n v_{\perp i}}$$

Mass flux q_m [kg/s] through cross section: $q_m = \bar{c} \cdot q_v$ where q_v [m³/s] gas flow rate.

$$q_v = \sum_{i=1}^n v_{\perp i} \Delta A_i$$

Measurement of concentration by sampling.



Sampling rig and sampling in a stack

- “CS” : probe’s head (in Hungarian “leszívó csonk”)
- “SZ” : probe’s tubing (in Hungarian “szonda szár”)
- “L” : particle separator (filter) (in Hungarian “leválasztó”)
- “M” : flow rate measurement (in Hungarian “terfo gatáram mérő”)
- “F” : blower with rpm regulation (in Hungarian “fűvő”)

\bar{c}_M : the average concentration from the measurement:

$$\bar{c}_M = \frac{\sum_{i=1}^n \frac{d_{szi}^2 \pi}{4} v_{szi} \Delta t_i c_{szi}}{\sum_{i=1}^n \frac{d_{szi}^2 \pi}{4} v_{szi} \Delta t_i}$$

In i-th sampling point:

v_{szi} sampling velocity

c_{szi} concentration in the sample probe

d_{szi} diameter of the probe’s head

Δt_i sampling time period.

By multiplying the numerator and denominator of the fraction by $v_{\perp i} / v_{\perp i}$:

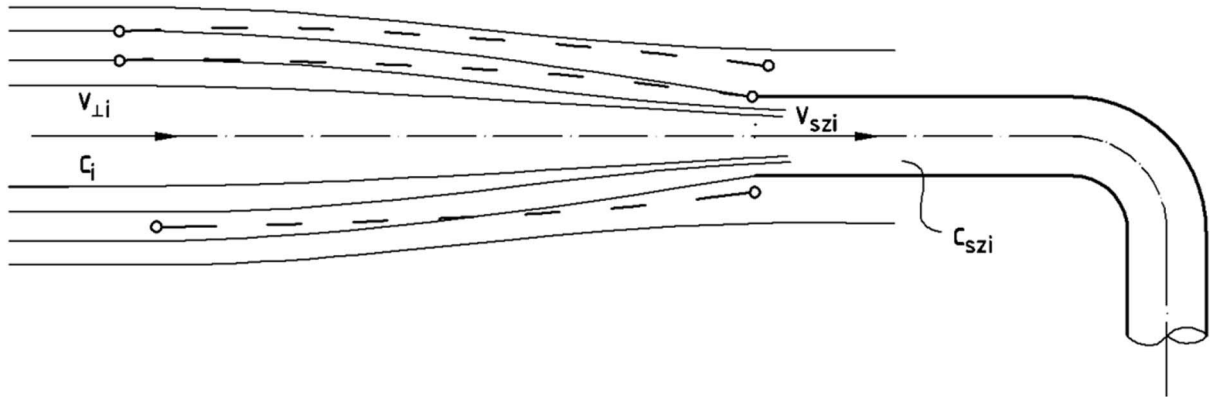
$$\bar{c}_M = \frac{\sum_{i=1}^n \frac{d_{szi}^2 \pi}{4} v_{szi} \Delta t_i c_{szi} \frac{v_{\perp i}}{v_{\perp i}}}{\sum_{i=1}^n \frac{d_{szi}^2 \pi}{4} v_{szi} \Delta t_i \frac{v_{\perp i}}{v_{\perp i}}}$$

If $\boxed{d_{szi}^2 \frac{v_{szi}}{v_{\perp i}} \Delta t_i \equiv \text{const.}}$,

$$\bar{c}_M \approx \frac{\sum_{i=1}^n c_{szi} v_{\perp i}}{\sum_{i=1}^n v_{\perp i}}. \text{ It is equal to } \boxed{\bar{c} \approx \frac{\sum_{i=1}^n c_i v_{\perp i}}{\sum_{i=1}^n v_{\perp i}}} \text{ if the concentration in sampling probe is equal to that of}$$

in front of the sampling head in the gas flow. $c_{szi} = c_i$.

Sampling velocity



Streamlines and paths of particles if $v_{szi} > v_{\perp i}$

The curved streamlines in front of the sampling probe cause differences in gas streamlines and paths of dust particles, so if $v_{szi} > v_{\perp i}$ the concentration in probe is smaller than that is in front of the probe.

The requirement of correct sampling procedure is: $\frac{v_{szi}}{v_{\perp i}} = 1$ (iso-kinetic sampling)

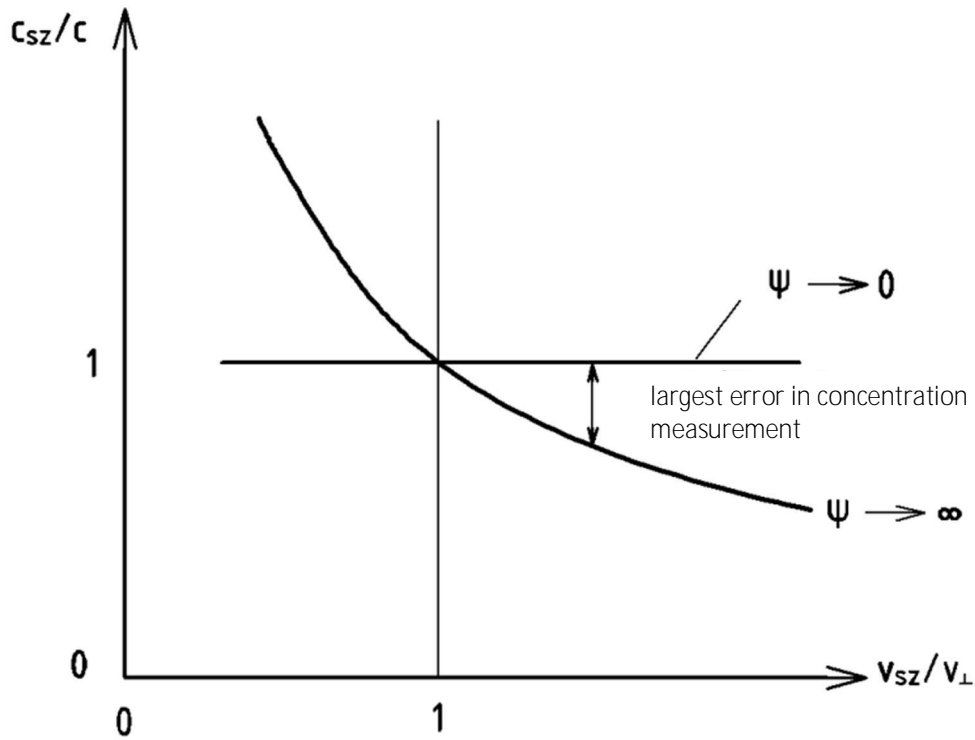
When iso-kinetic sampling is applied: $c_{szi} = c_i$.

In case of deviation from the iso-kinetic sampling the largest error (at $\psi \rightarrow \infty$) in concentration measurement:

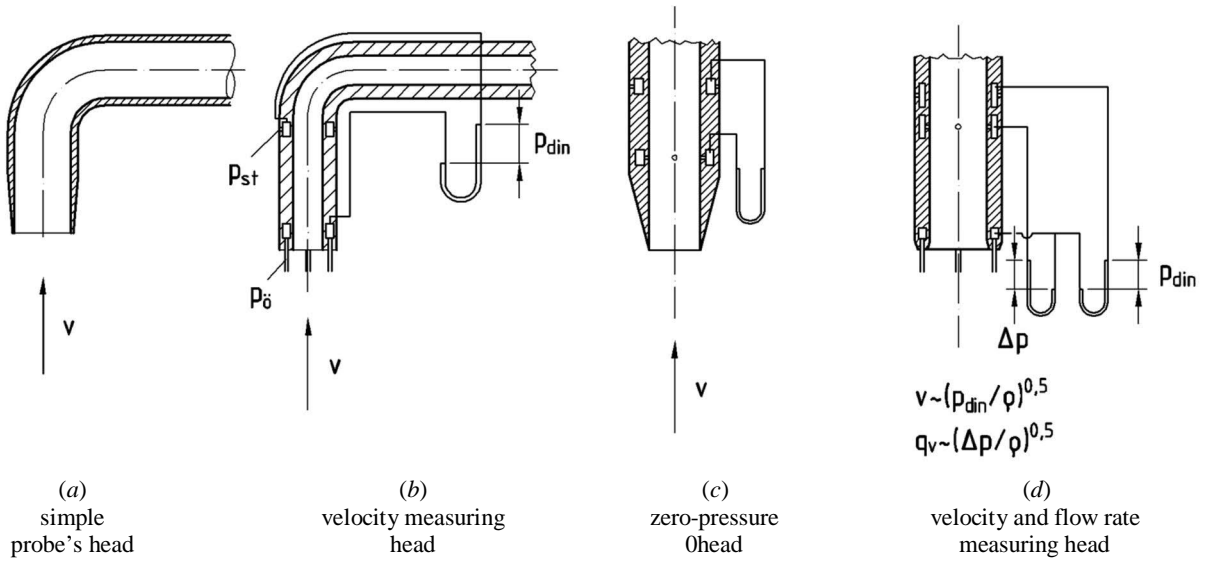
$$c_{sz} = \frac{v_{\perp} \frac{d_{sz}^2 \pi}{4} c}{v_{sz} \frac{d_{sz}^2 \pi}{4}} = \frac{v_{\perp} c}{v_{sz}} \Rightarrow \frac{c_{sz}}{c} = \frac{1}{v_{sz} / v_{\perp}}.$$

Problems of sampling:

- deviation from iso-kinetic sampling (measuring local velocity and sample flow rate, pressures, temperatures),
- changing operation state of the plant,
- deposition of the dust in the probe's head of the sampling apparatus,
- cooling of sampled gas,
- condensation of the humidity in the probe's head (can be solved by heating, particle separator in the stack).



Types of sampling probes: simple, probe capable for measurement of local velocity, “zero pressure” type probe’s head, probe capable for measuring both velocity and flow rate.



Mass balance (mass flow rate balance) of a separator

M_{in} [kg/s] mass flow rate of pollutants entering the separator (“in”)

M_{sep} [kg/s] mass flow rate of pollutants separated (“sep”)

M_{out} [kg/s] mass flow rate of pollutants escaping the separator (“out”)

Balance: $M_{in} = M_{out} + M_{ssep}$

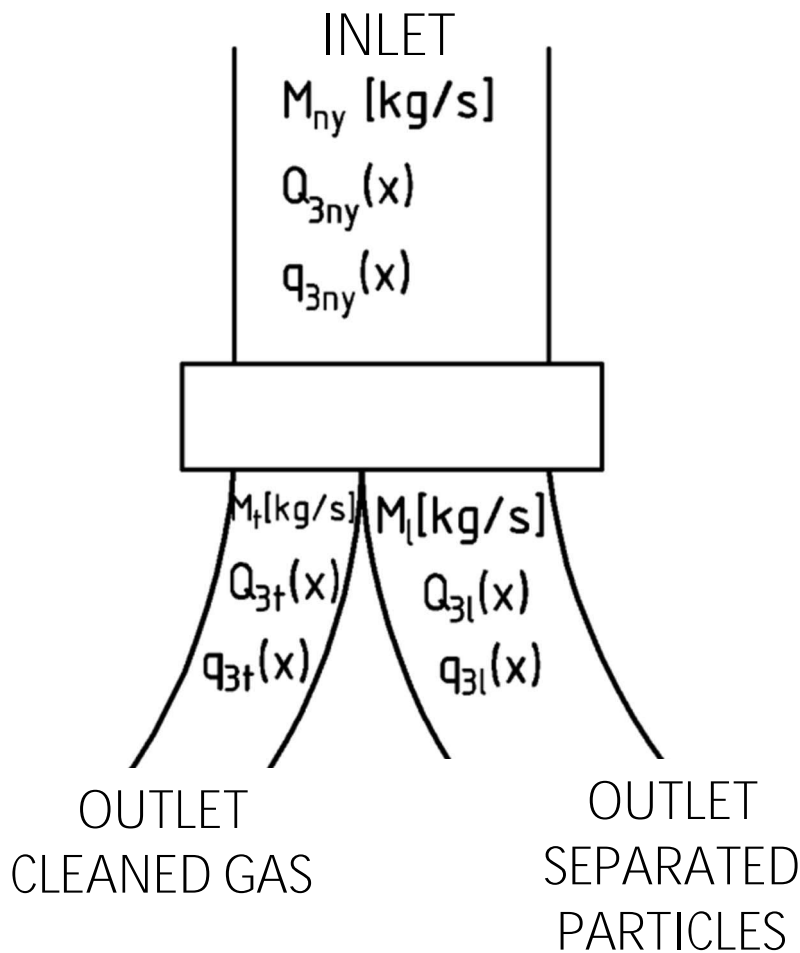
$$1 = \frac{M_{sep}}{M_{in}} + \frac{M_{out}}{M_{in}} = E + P,$$

where $E = M_{sep}/M_{in}$ overall efficiency of separation, and $P = M_{out}/M_{in}$ penetration

$$M_{in} = q_v c_{in}$$

$$M_{out} = q_v c_{out}$$

$$M_{sep} = q_v (c_{in} - c_{out})$$

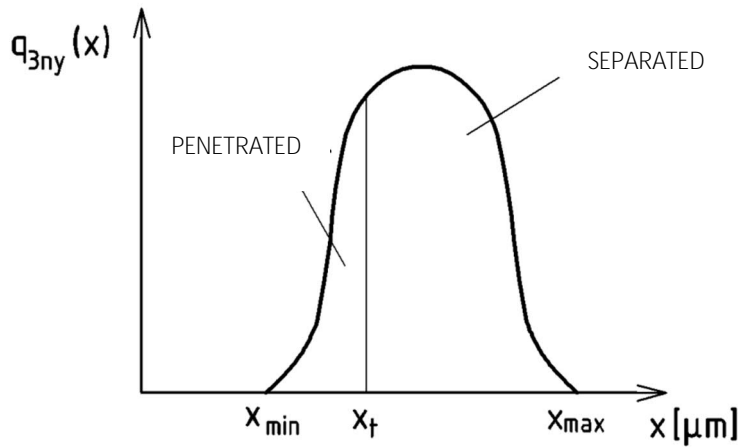


Overall efficiency expressed by c [kg/m³] pollutant concentration:

$$E = \frac{M_s}{M_{in}} = \frac{c_{in} - c_{out}}{c_{in}} = 1 - \frac{c_{out}}{c_{in}}.$$

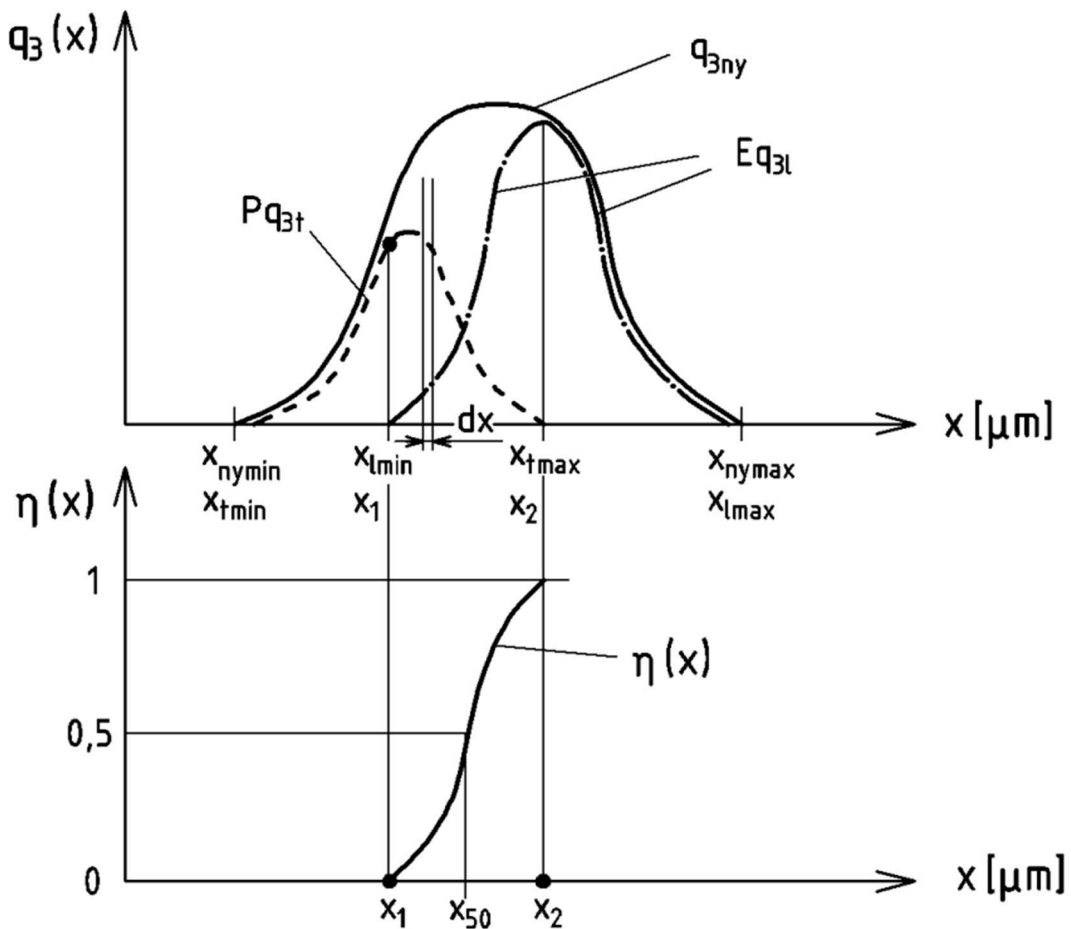
Fractional efficiency (size dependent) $\eta(x)$

- a) idealistic separation means, that there is a limiting diameter (x_t), and
 $x < x_t$: all particles in this size range are penetrated ($\eta(x)=0\%$)
 $x_t < x$: all particles in this size range are separated ($\eta(x)=100\%$)



$$E = \int_{x_t}^{x_{max}} q_3(x) dx, \quad P = \int_{x_{min}}^{x_t} q_3(x) dx, \quad E + P = 1$$

- b) In a real separation process there is no limiting diameter but a $x_1 - x_2$ size range



- $x < x_1$: all particles in this size range are penetrated ($\eta(x)=0\%$)
 $x_1 < x < x_2$: as diameter increases the fractional efficiency increases
 $x_2 < x$: all particles in this size range are separated ($\eta(x)=100\%$)

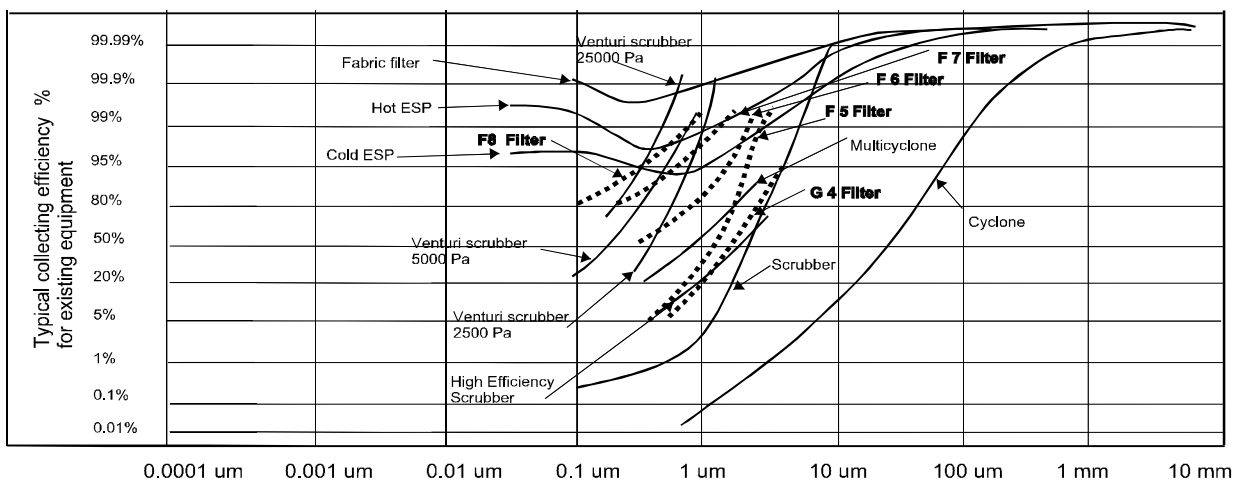
Knowing the mass flow rates (M) and q density functions in the incoming (“in”), outgoing (“out”) and separated (“sep”) particle phases we get:

$$M_{in} q_{3,in} dx = M_{out} q_{3,out} dx + M_{sep} q_{3,sep} dx$$

$$q_{3,in} = \frac{M_{out}}{M_{in}} q_{3,out} + \frac{M_{sep}}{M_{in}} q_{3,sep} = P q_{3,out} + E q_{3,sep}$$

Fractional efficiency can be calculated:

$$\eta(x) = \frac{dM_{sep}(x)}{dM_{in}(x)} = \frac{M_{sep} q_{3,sep}(x) dx}{M_{in} q_{3,in}(x) dx} = \frac{E q_{3,sep}}{q_{3,in}}$$



Two steps of separation process

1st step:

The particles should be removed from the gas streamline: they should be moved relative to the gas

in order to establish

- contact with solid surfaces,
- contact fluid films, of drops much bigger than the particles,
- contact with other particles

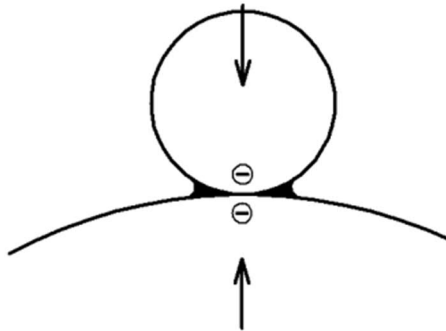
by utilizing effects listed below that allow for moving the particles relative to the gas:

- inertia of particle
- gravitational force
- diffusion caused by thermal agitation of gas molecules
- additional external (e.g. electrostatic) forces caused by charging of particles

2nd step:

Forces should be utilized to “stick” the particles to each other, to solid surfaces and to water films. Such forces are:

- Van der Waals force: is the attractive force between molecules
- Electrostatic attraction
- Surface tension, a property of the surface of a liquid that causes it to behave as an elastic sheet



Particles “sticks” due to the surface tension of water bridge.

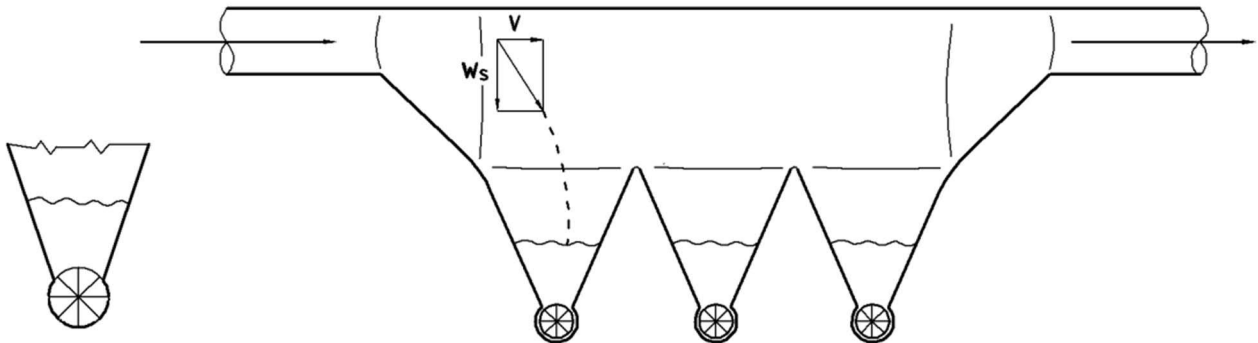
As result of the above 1&2 steps:

- Conglomeration of particles or droplets will have higher settling velocity (and inertia parameter) than a single particle, hence can be separated with higher efficiency.
- Particles adhere to the filter material (to a solid surface), and they can be removed together with the filter.
- Particles are collected by large water droplets that can be collected in an easier way, or particles are collected by using water films on the solid surfaces.

4. Gas cleaning equipments, various types of separators

Settling chambers

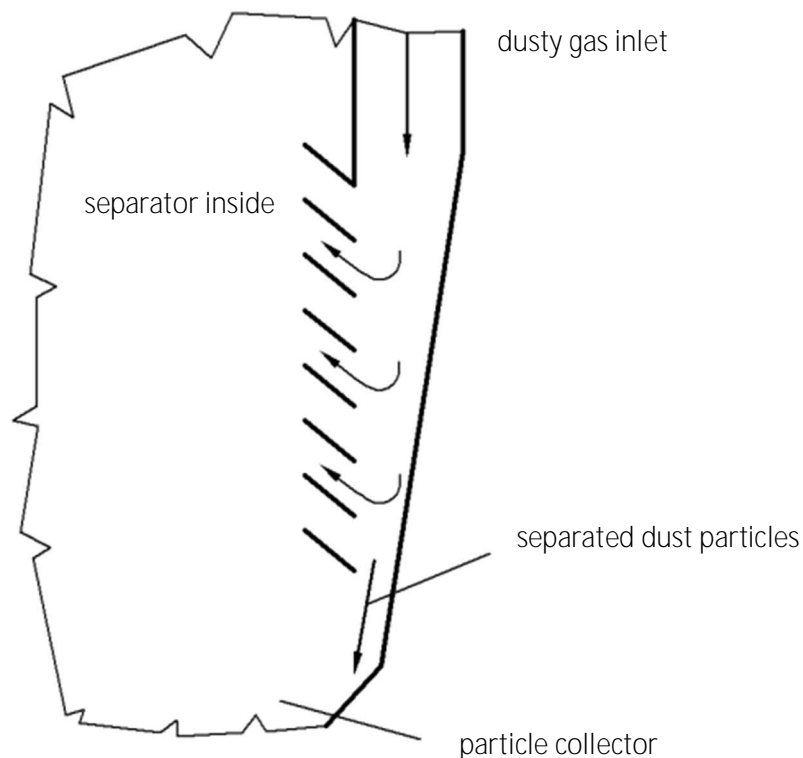
The particles are moved relative to the gas by gravitational force and Van der Waals force is utilized to “stick” the particles to each other in the hopper.



In general the pressure in the separators is lower than the ambient, so at the removal of dust the air flow of the hopper is not allowed. The separated dust is removed from the hopper by a charging screw or rotating cells charger.

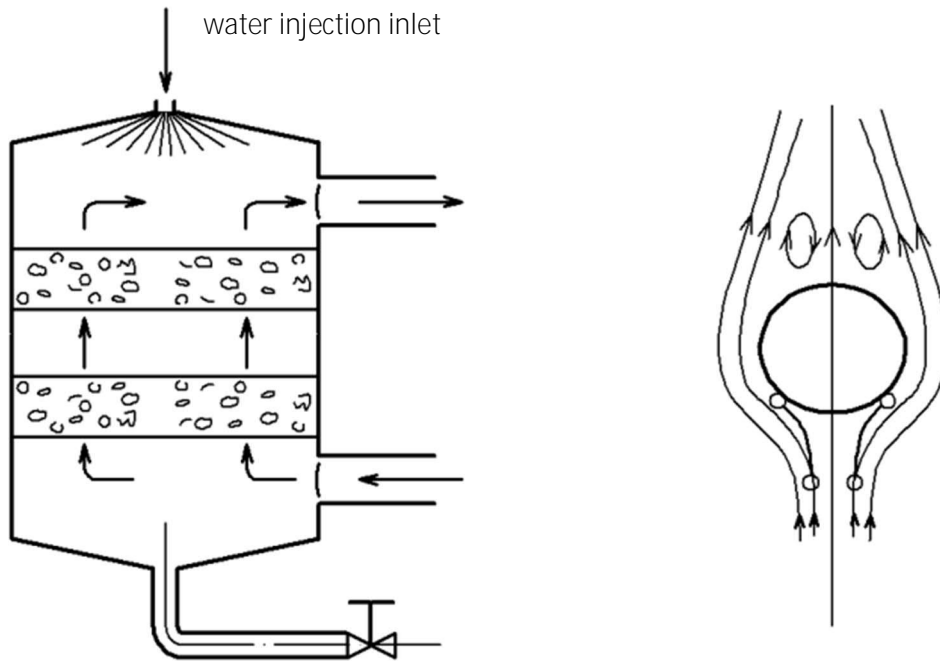
Pre-separator louvers

In pre-separator louvers the particles are moved relative to the gas by their inertia as well as gravitational force and Van der Waals force is utilized to “stick” the particles to each other in the hopper.



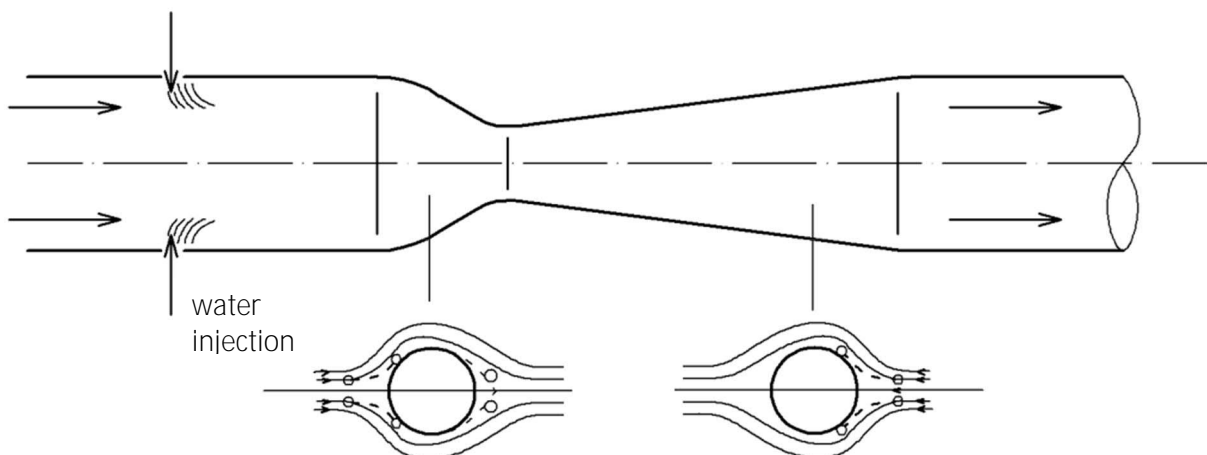
Scrubbers (washing towers)

The particles are moved relative to the gas and reach the relatively large water droplets mainly by their inertia as well as diffusion and the surface tension is utilized to “stick” the particles to the droplets.



Venturi scrubber

The particles are moved relative to the gas and reach the relatively large water droplets mainly by their inertia as well as diffusion and the surface tension is utilized to “stick” the particles to the droplets. The large droplets are separated by a cyclone separator.



Cyclones

The particles are moved relative to the gas by their inertia to the wall and move down in a spiral path to the hopper by gravitational force and Van der Waals force is utilized to “stick” the particles to each other in the hopper.

Simple considerations results in estimation of the limit particle diameter (x_{limit}): diameter of smallest particle separated by a cyclone. Equilibrium of the forces acting on a particle moving along a circular orbit of radius r_1 with tangential velocity v_{1t} . In order to maintain this motion a centripetal force is needed acting on the particle:

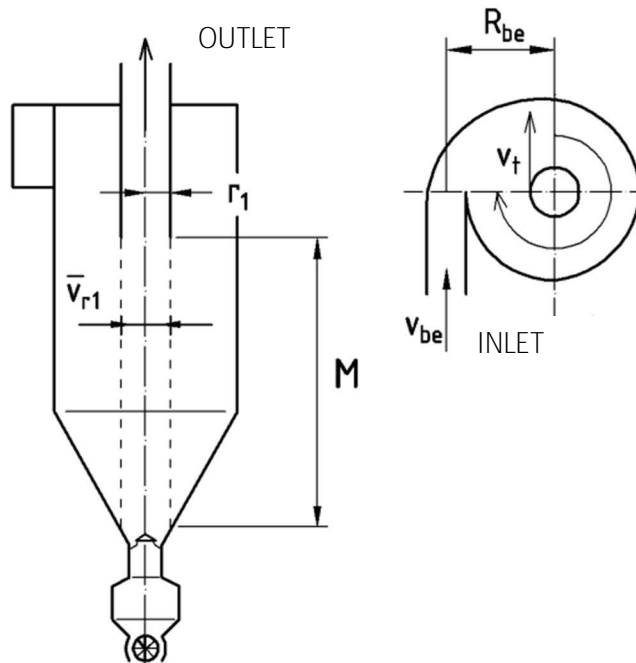
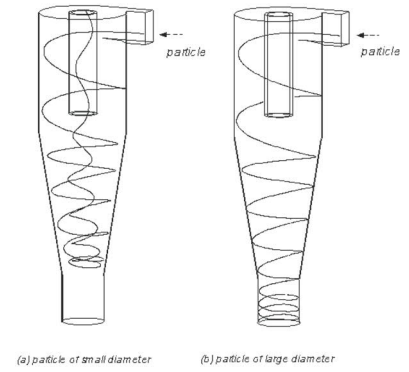
$$F_c = \frac{x_{limit}^3 \pi}{6} \rho_p \frac{v_{1t}^2}{r_1}.$$

Here the radial velocity of gas is

$$v_{1r}: \frac{x_{limit}^3 \pi}{6} \rho_p \frac{v_{1t}^2}{r_1} = 3\pi \mu x_{limit} v_{1r},$$

where

$$\bar{v}_{1r} \cong \frac{q_v}{2r_1 \pi M}.$$



Tangential velocity can be calculated by assuming a potential vortex in the cyclone:

$$v_{1t} = \frac{R_{in} v_{in}}{r_1}, \text{ where } R_{in} \text{ is the radius of air tangential introduction of polluted gas into the cyclone,}$$

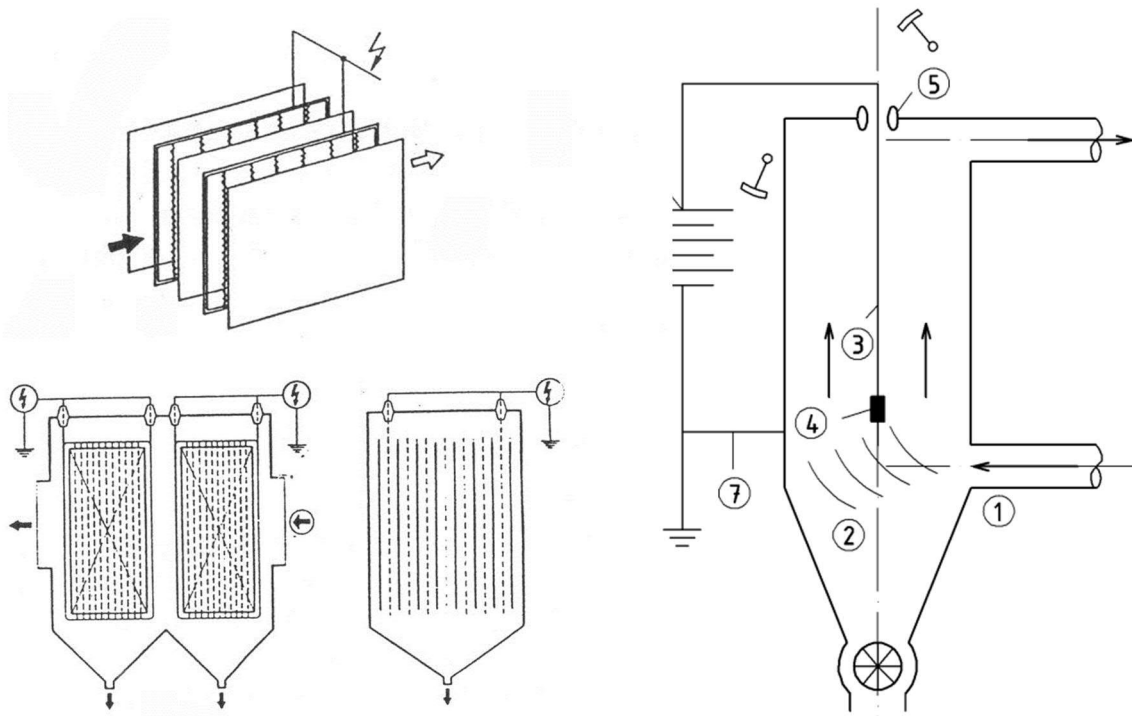
with inlet velocity v_{in} . By expressing v_{1r} and substituting it we get:

$$\frac{v_{in}^2 R_{in}^2}{r_1 r_1^2} \frac{x_{limit}^2}{18 \mu} \rho_p = v_{1r} = \frac{q_v}{2r_1 \pi M}.$$

The expression for limit particle diameter (x_{limit}) is:
$$x_{limit} = \sqrt{\frac{9}{\pi} \frac{r_1}{R_{in} v_{in}} \sqrt{\frac{\mu q_v}{\rho_p M}}}$$

Electrostatic precipitators (ESP)

Venturi scrubbers, electrostatic precipitators and fibrous filters meet the most rigorous environmental protection requirements.



1. Inflow of dusty gas
2. Turning vanes, perforated plates for providing uniform flow
3. High voltage electrode
4. Weight spanning the electrode
5. Insulator
6. High voltage direct current
7. Grounding, collecting electrode

At 10-30 kV voltage corona discharge – electrons moving in direction of collecting electrode and ionize gas molecules, producing further electrons (electron avalanche). Electron-affine gas molecules (O_2 , SO_2) capture electrons: molecules of negative charge move in the electrostatic field of force. They load on particles – so particles will be charged, and move towards collecting electrodes.

The charged particles are moved relative to the gas by the electrostatic field of force to the collecting electrode and Van der Waals and electrostatic forces are utilized to “stick” the particles to each other. The particles are removed from the collecting electrodes to the hopper by keeping periodically the collecting electrodes hitting by hammers. The uniformity of flow and low turbulence is crucial.

Problems: instable operation in case of particles of small electric conductivity (high electric resistance), leakage, considerable reduction of efficiency, difficulties of cleaning of collecting electrodes, in case of electric breakdown voltage of corona the particles can escape the separator easily.

Filters (did not presented, hence it is just for your information)

Filters: porous sheets (woven fabrics or flises) of elementary fibres (filter mats).

Example: mass of 1 m² filter: m_f = 0.4 kg/m², thickness: s = 0.004 m, diameter and density of fibres d_f = 20 μm and ρ_f = 1000 kg/m³.

Solidity: $\alpha = \frac{V_f}{V} = 1 - \varepsilon$

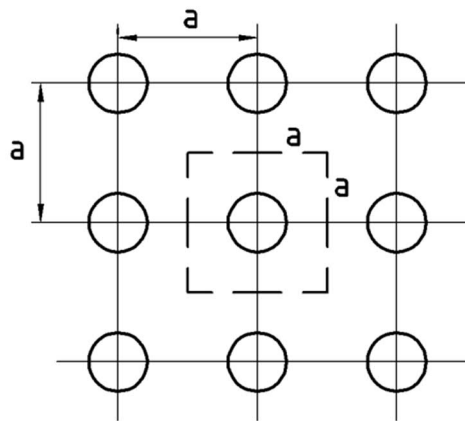
V_f and V[m³] volume of fibres and filter, ε porosity. $\alpha = \frac{m_f}{\rho_f s} = 0.1$. Usual values: α = 0,01 ÷ 0,3

Length of the fibres in 1 m² filter sheet: $L_{f1m^2} = \frac{V_f}{d^2 \pi / 4} = 1.27 \cdot 10^6 m / m^2$,

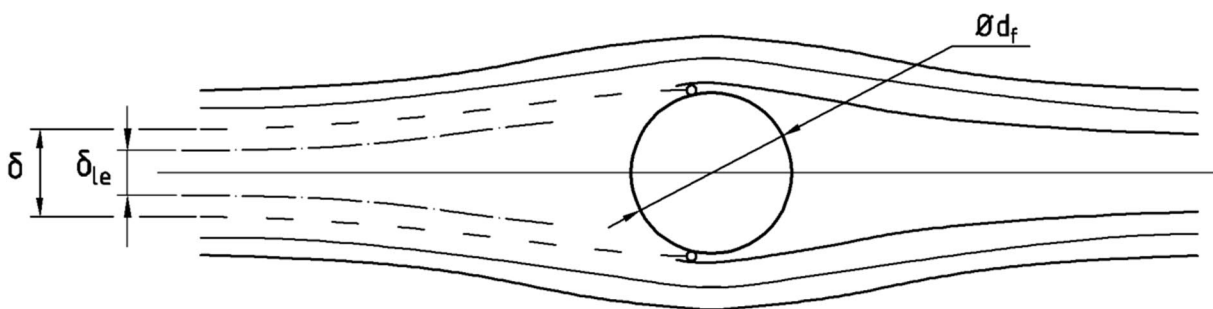
Surface area: $A_{f1m^2} = L_{f1m^2} d \pi = 80 m^2 / m^2$. Average distance between fibres

$\alpha = \frac{d_f^2 \pi}{4a^2} \Rightarrow \frac{a}{d_f} = \sqrt{\frac{\pi}{4\alpha}} = 2.8$.

If d_f = 20 μm, a = 60 μm, in s = 4 mm 70 rows of fibres.

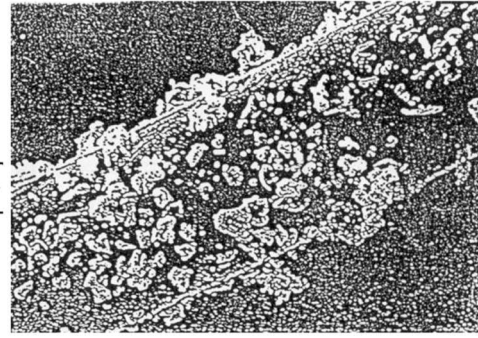
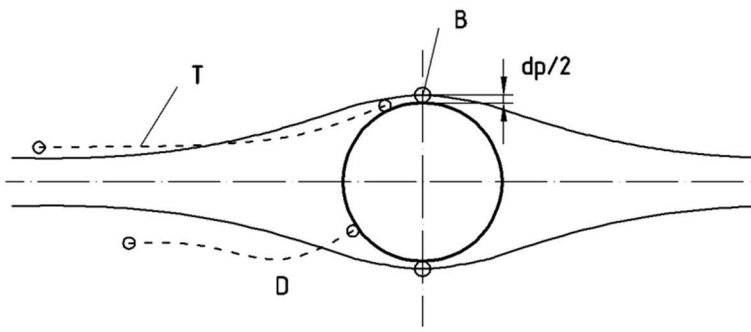


The separated particles are of diameter of several microns, so the particles are not filtered in a conventional way. The collection of particles occurs in another way: Δp pressure difference across the filter induces flow of dusty gas across the filter mat: v_f [m/s] filtration velocity. In the flow past fibres viscosity is the dominant.



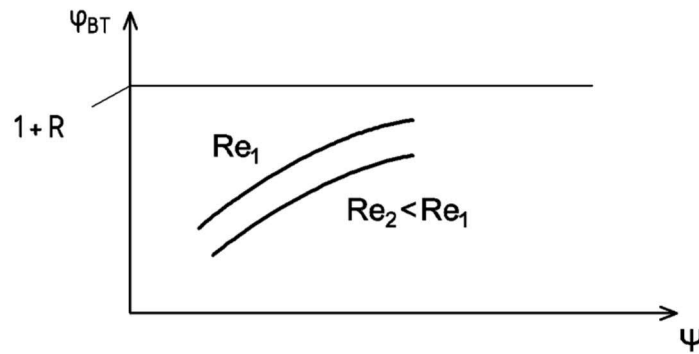
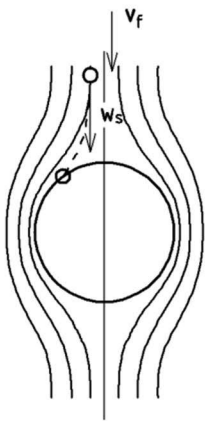
There are a number of mechanisms moving the particles relative to the gas towards the fibres and causes collision of particles on the fibre:.

- curved streamlines: inertia of particles
- weight
- diffusion
- electrostatic forces
- blocking



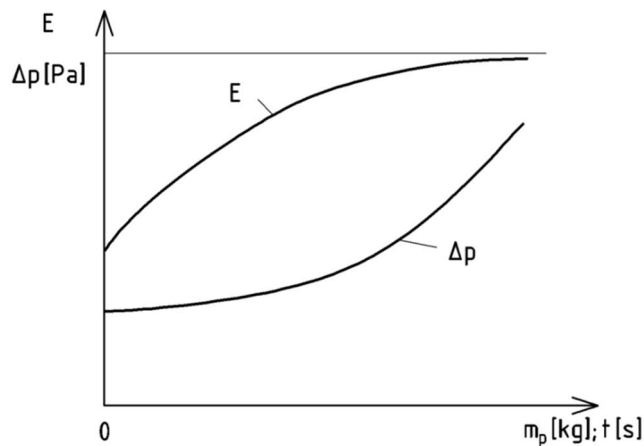
T inertia, B blocking, D diffusion

Collision efficiency of single fibre: $\phi = \frac{\delta}{d_f}$, where δ is the thickness of layer of particles colliding the fibre. Part of the particles is collected by fibres and other part rebounds. Van der Waals and electrostatic forces as well as surface tension are utilized to "stick" the particles to the fibres and to each other.



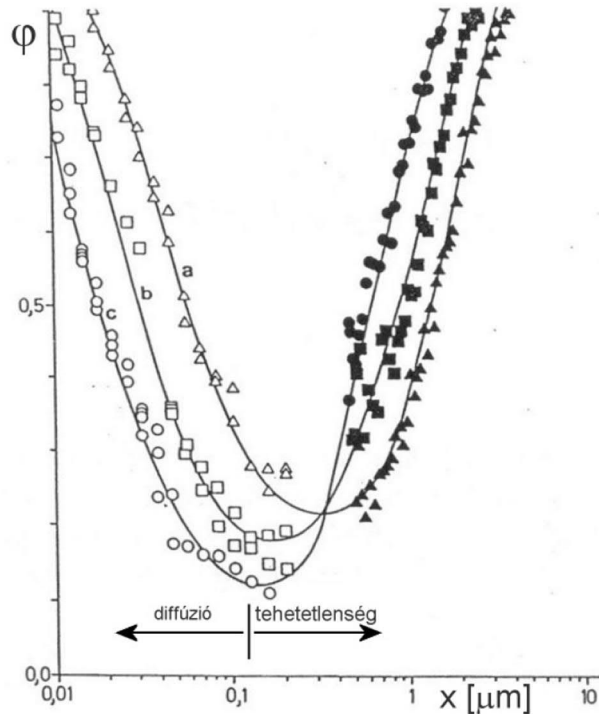
Collision efficiency as function of ψ inertia parameter: $\Psi = \frac{v_f w_s}{g d_f} = \frac{v_f \rho_p x^2}{18\mu d_f}$

Collected particles increase the collecting surface and the solidity (layer of particle and dendrites) so both the efficiency and the pressure drop increase in time. The deposited particles can be removed from the *deep bed filters* so they should be replaced if the pressure drop exceeds a given value. So they can be used only for small concentration (e.g. aftercleaners or at ventilation of clean rooms).



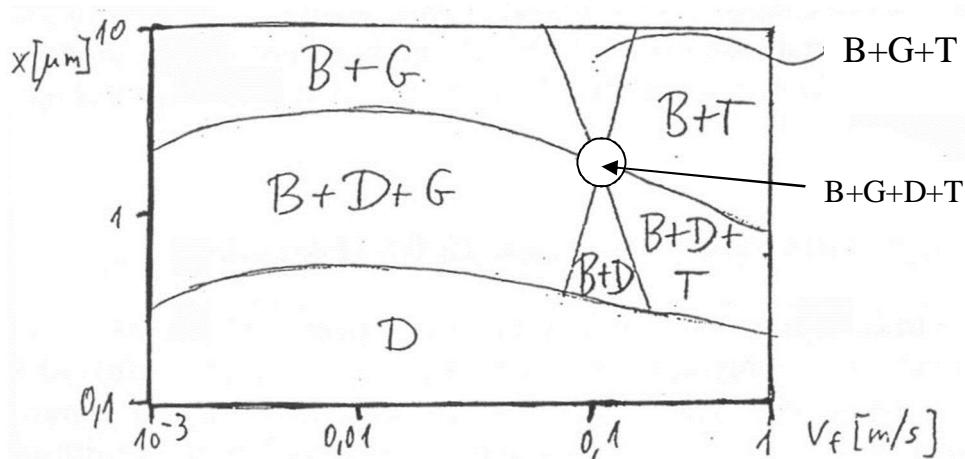
Change of pressure drop and filtration efficiency in time (separated dust mass)

The most particles are deposited in the layer of the filter mat close to its raw gas side. If the solidity of filter is higher, a layer of particles (dust cake) develops on the surface of the filter that separates the particles. This dust cake can be removed by shaking, deforming the filter: regeneration of filters. Polluted gases of much higher concentration than at deep bed filters can be cleaned by this type of filters: *surface filters*.



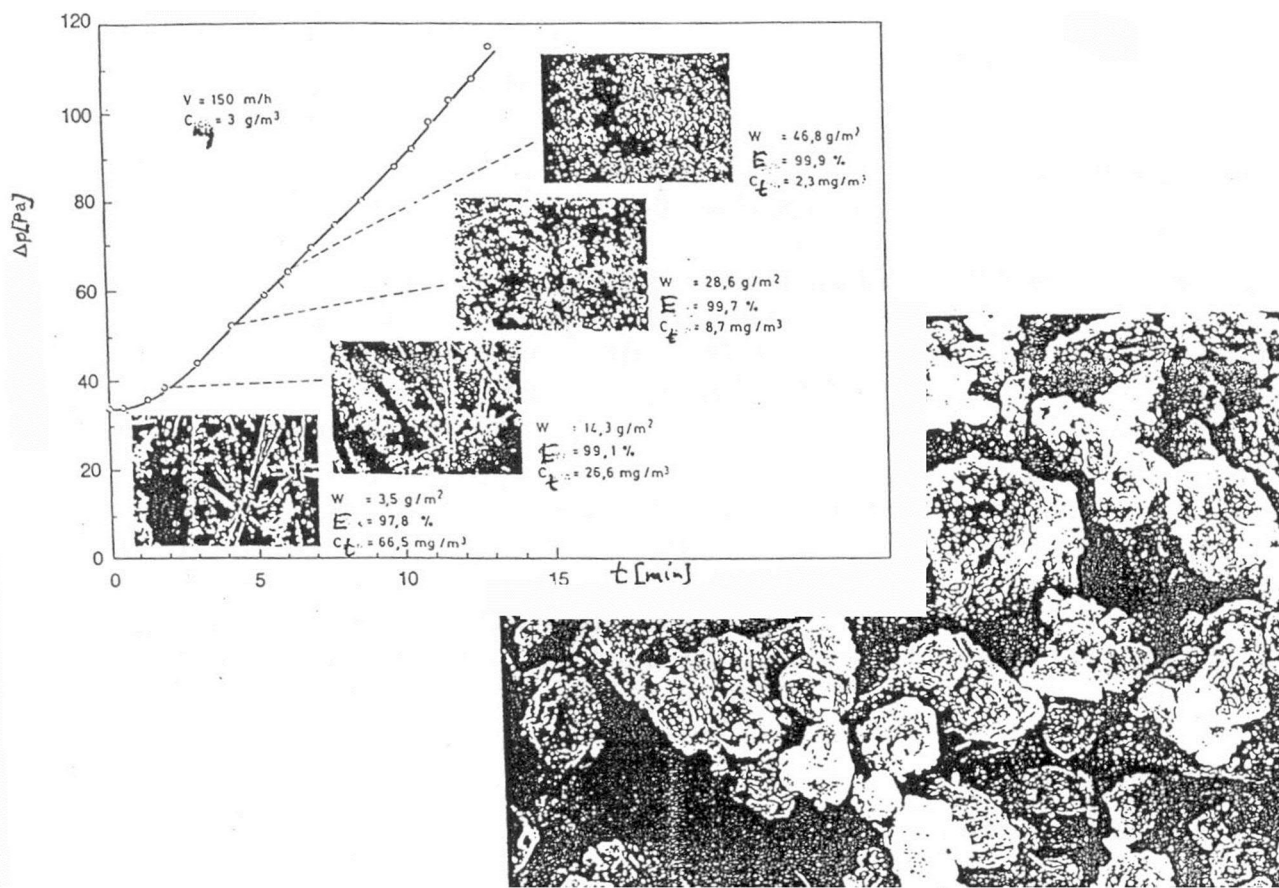
Collision efficiency as function of particle diameter and filtration velocity
 $v_c > v_b > v_a$

The domain of dominant effects influencing the collision efficiency



Where: T inertia, G weight, D diffusion, B blocking

Surface filters



Stable and instable operation of the bag filter unit:

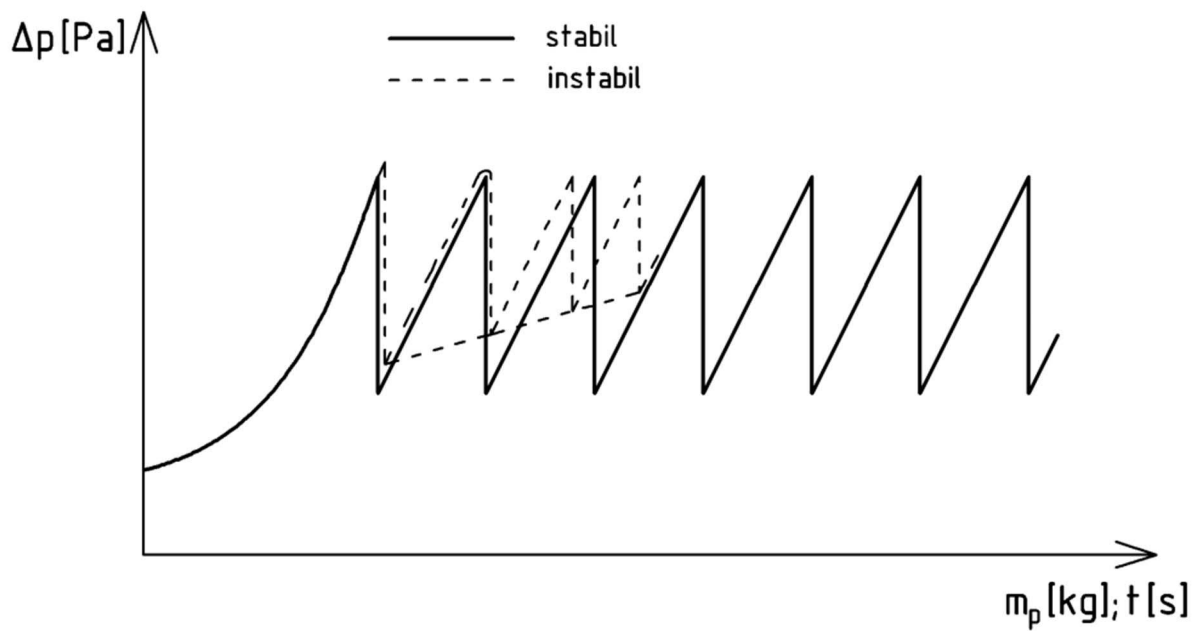


TABLE 25-13

Advantages and Disadvantages of Cyclone Collectors

(SOURCE: "Perry's Chemical Engineers' Handbook" ed. Don W. Green | 1997 | ISBN 0-07-049841-5 | McGraw-Hill)

Advantages

1. Low cost of construction
2. Relatively simple equipment with few maintenance problems
3. Relatively low operating pressure drops (for degree of particulate removal obtained) in the range of approximately 2- to 6-in water column
4. Temperature and pressure limitations imposed only by the materials of construction used
5. Collected material recovered dry for subsequent processing or disposal
6. Relatively small space requirements

Disadvantages

1. Relatively low overall particulate collection efficiencies, especially on particulates below 10 μm in size
2. Inability to handle tacky materials

TABLE 25-14

Advantages and Disadvantages of Wet Scrubbers

(SOURCE: "Perry's Chemical Engineers' Handbook" ed. Don W. Green | 1997 | ISBN 0-07-049841-5 | McGraw-Hill)

Advantages

1. No secondary dust sources
2. Relatively small space requirements
3. Ability to collect gases as well as particulates (especially "sticky" ones)
4. Ability to handle high-temperature, high-humidity gas streams
5. Capital cost low (if wastewater treatment system not required)
6. For some processes, gas stream already at high pressures (so pressure drop considerations may not be significant)
7. Ability to achieve high collection efficiencies on fine particulates (however, at the expense of pressure drop)
8. Ability to handle gas streams containing flammable or explosive materials

Disadvantages

1. Possible creation of water-disposal problem
2. Product collected wet
3. Corrosion problems more severe than with dry systems
4. Steam plume opacity and/or droplet entrainment possibly objectionable
5. Pressure-drop and horsepower requirements possibly high
6. Solids buildup at the wet-dry interface possibly a problem
7. Relatively high maintenance costs
8. Must be protected from freezing
9. Low exit gas temperature reduces exhaust plume dispersion
10. Moist exhaust gas precludes use of most additional controls

TABLE 25-15

Advantages and Disadvantages of Dry Scrubbers

(SOURCE: "Perry's Chemical Engineers' Handbook" ed. Don W. Green | 1997 | ISBN 0-07-049841-5 | McGraw-Hill)

Advantages

1. No wet sludge to dispose of
2. Relatively small space requirements
3. Ability to collect acid gases at high efficiencies
4. Ability to handle high-temperature gas streams
5. Dry exhaust allows addition of fabric filter to control particulate

Disadvantages

1. Acid gas control efficiency not as high as with wet scrubber
2. No particulate collection—dry scrubber generates particulate
3. Corrosion problems more severe than with dry systems
4. Solids buildup at the wet-dry interface possibly a problem
5. Relatively high maintenance costs
6. Must be protected from freezing
7. Low exit gas temperature reduces exhaust plume dispersion

TABLE 25-16

Advantages and Disadvantages of Electrostatic Precipitators

(SOURCE: "Perry's Chemical Engineers' Handbook" ed. Don W. Green | 1997 | ISBN 0-07-049841-5 | McGraw-Hill)

Advantages

1. Extremely high particulate (coarse and fine) collection efficiencies attainable (at a relatively low expenditure of energy)
2. Collected material recovered dry for subsequent processing or disposal
3. Low pressure drop
4. Designed for continuous operation with minimum maintenance requirements
5. Relatively low operating costs
6. Capable of operation under high pressure (to 150 lbf/in²) or vacuum conditions
7. Capable of operation at high temperatures [to 704 °C(1300 °F)]
8. Relatively large gas flow rates capable of effective handling

Disadvantages

1. High capital cost
2. Very sensitive to fluctuations in gas-stream conditions (in particular, flows, temperature, particulate and gas composition, and particulate loadings)
3. Certain particulates difficult to collect owing to extremely high- or lowresistivity characteristics
4. Relatively large space requirements required for installation
5. Explosion hazard when treating combustible gases and/or collecting combustible particulates
6. Special precautions required to safeguard personnel from the high voltage
7. Ozone produced by the negatively charged discharge electrode during gas ionization

8. Relatively sophisticated maintenance personnel required
9. Gas ionization may cause dissociation of gas stream constituents and result in creation of toxic by-products
10. Sticky particulates may be difficult to remove from plates
11. Not effective in capturing some contaminants that exist as vapours at high temperatures (e.g., heavy metals, dioxins)

TABLE 25-17

Advantages and Disadvantages of Fabric-Filter Systems

(SOURCE: "Perry's Chemical Engineers' Handbook" ed. Don W. Green | 1997 | ISBN 0-07-049841-5 | McGraw-Hill)

Advantages

1. Extremely high collection efficiency on both coarse and fine (sub-micrometer) particles
2. Relatively insensitive to gas-stream fluctuation; efficiency and pressure drop relatively unaffected by large changes in inlet dust loadings for continuously cleaned filters
3. Filter outlet air capable of being recirculated within the plant in many cases (for energy conservation)
4. Collected material recovered dry for subsequent processing or disposal
5. No problems with liquid-waste disposal, water pollution, or liquid freezing
6. Corrosion and rusting of components usually not problems
7. No hazard of high voltage, simplifying maintenance and repair and permitting collection of flammable dusts
8. Use of selected fibrous or granular filter aids (precoating), permitting the high-efficiency collection of submicrometer smokes and gaseous contaminants
9. Filter collectors available in a large number of configurations, resulting in a range of dimensions and inlet and outlet flange locations to suit installment requirements
10. Relatively simple operation

Disadvantages

1. Temperatures much in excess of 288 °C (550 °F) requiring special refractory mineral or metallic fabrics that are still in the developmental stage and can be very expensive
2. Certain dusts possibly requiring fabric treatments to reduce dust seeping or, in other cases, assist in the removal of the collected dust
3. Concentrations of some dusts in the collector (~50 g/m³) forming a possible fire or explosion hazard if a spark or flame is admitted by accident; possibility of fabrics burning if readily oxidizable dust is being collected
4. Relatively high maintenance requirements (bag replacement, etc.)
5. Fabric life possibly shortened at elevated temperatures and in the presence of acid or alkaline particulate or gas constituents
6. Hygroscopic materials, condensation of moisture, or tarry adhesive components possibly causing crusty caking or plugging of the fabric or requiring special additives
7. Replacement of fabric, possibly requiring respiratory protection for maintenance personnel
8. Medium pressure-drop requirements, typically in the range 4- to 10-in water column

TABLE 25-18

Advantages and Disadvantages of Absorption Systems (Packed and Plate Columns)

(SOURCE: "Perry's Chemical Engineers' Handbook" ed. Don W. Green | 1997 | ISBN 0-07-049841-5 | McGraw-Hill)

Advantages

1. Relatively low pressure drop
2. Standardization in fiberglass-reinforced plastic (FRP) construction permitting operation in highly corrosive atmospheres
3. Capable of achieving relatively high mass-transfer efficiencies
4. Increasing the height and/or type of packing or number of plates capable of improving mass transfer without purchasing a new piece of equipment
5. Relatively low capital cost
6. Relatively small space requirements
7. Ability to collect particulates as well as gases
8. Collected substances may be recovered by distillation

Disadvantages

1. Possibility of creating water (or liquid) disposal problem
2. Product collected wet
3. Particulates deposition possibly causing plugging of the bed or plates
4. When FRP construction is used, sensitive to temperature
5. Relatively high maintenance costs
6. Must be protected from freezing

TABLE 25-19

Comparison of Plate and Packed Columns

(SOURCE: "Perry's Chemical Engineers' Handbook" ed. Don W. Green | 1997 | ISBN 0-07-049841-5 | McGraw-Hill)

Packed column

1. Lower pressure drop
2. Simpler and cheaper to construct
3. Preferable for liquids with high-foaming tendencies

Plate column

1. Less susceptible to plugging
2. Less weight
3. Less of a problem with channeling
4. Temperature surge resulting in less damage

4th PART

Questions to the 4th test

Similar questions will be given in these topics. Test questions are based on the oral lectures, presented slides and this lecture handout.

1. Definition of aerosols. Types of aerosols. Evaluation & Conclusions.
2. Equivalent diameter of non-spherical particles. Types of equivalences. Evaluation & Conclusions.
3. Characterisation of particle-laden flows. Average relative distance between particles in gas. Volume ratio and mass loading ratio. Evaluation & Conclusions.
4. Characterization of particle assembly (Q_0, q_0) Evaluation & Conclusions.
5. Navier-Stokes equation including effect of particles. Conditions for neglectation of the effect of particles on the gas flow field. Evaluation & Conclusions.
6. Aerodynamic drag force (Stokes' drag form) acting on a single particle, drag coefficient. Evaluation & Conclusions.
7. Momentum equation for particles in gas flow. Evaluation & Conclusions.
8. Calculation of settling velocity. Evaluation & Conclusions.
9. Determination of particle path based on the inertia parameter (Ψ_p) / particle Stokes number (St_p) in case of a curved gas streamline of carrier flow. Evaluation & Conclusions.
10. Mass balance of a separator, overall efficiency, penetration. Ideal / real separation: fractional efficiency. Evaluation & Conclusions.
11. Main conditions of proper sampling during the measurement of particle concentration in a duct. Evaluation & Conclusions.
12. Steps of separation, forces and effects acting on the particles, relative to the gas & forces adhering the particles to each other, to solid surfaces and to water droplets of films. Evaluation & Conclusions.
13. Settling chambers / pre-separator louvers / washing tower / Venturi scrubber / cyclone / electrostatic precipitator / filters (one of your choice): sketch of the facility, operational basics, main forces & effects in the separation process, efficiency, costs etc.).