Lateral dispersion of fuel in a stationary fluidised bed boiler

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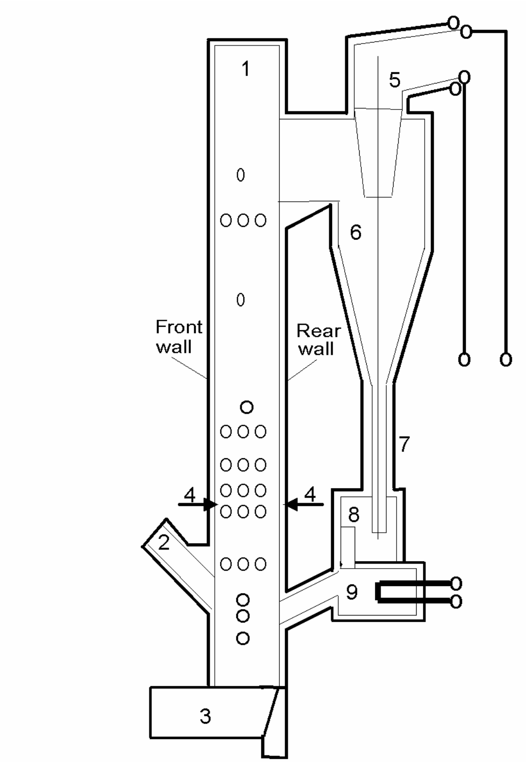
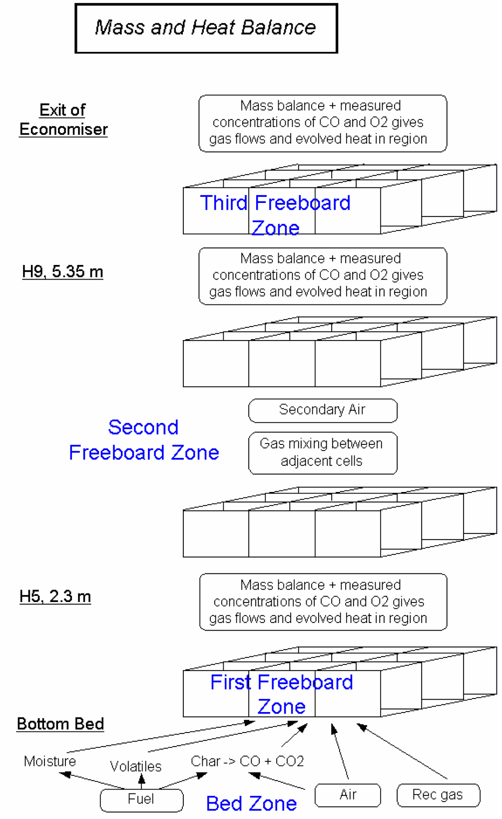
Heat balance modelling



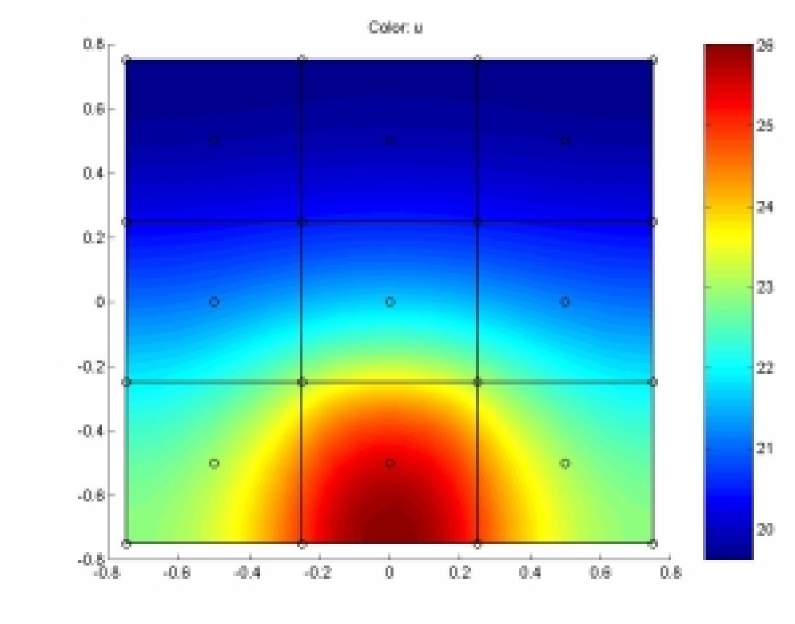
The combustion chamber is divided in to a cell-structure



Fuel dispersion – concentration

- **The rates of devolatilisation and dispersion determines the lateral distribution of the volatiles released in the bottom bed.**

$$-\nabla \cdot (D_{sr} \nabla C) + \frac{C}{\tau} = S$$



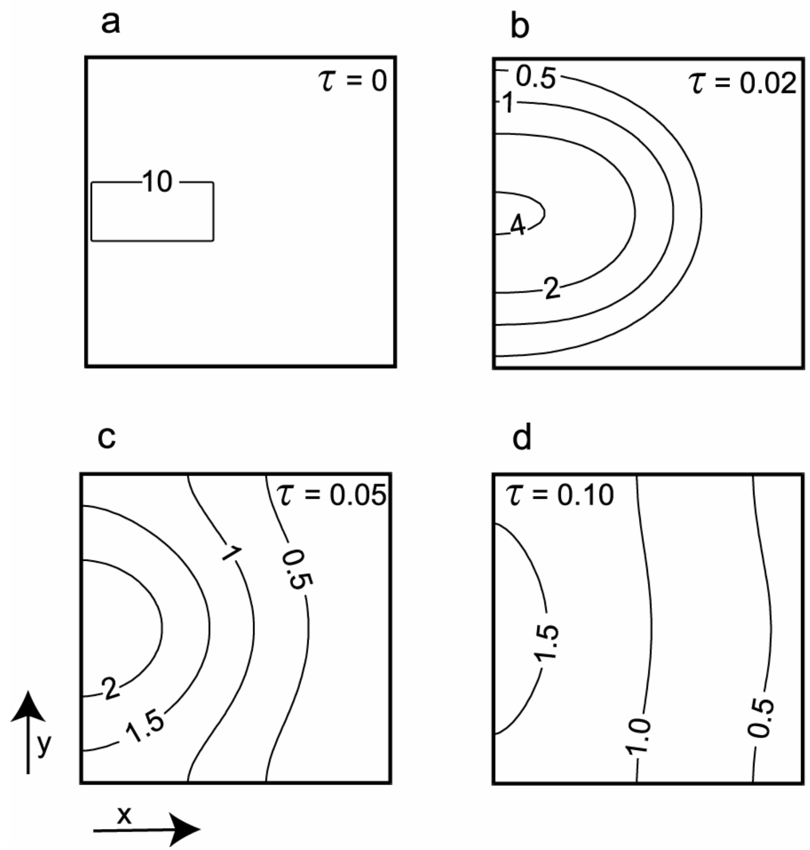
C - Volatiles in the fuel
D_{sr} - Dispersion coefficient
τ - Characteristic devolatilisation time
S - Source term

Lateral Tracer Dispersion

Solution of the time-dependent dispersion equation:

$$\frac{\partial C}{\partial t} = D_{sr} \cdot \nabla^2 C$$

The lateral tracer concentration is the integrated value along the height of the bed



Circulation pattern

The transport of solids could be divided into two processes; dispersion (random) and convection

The convective circulation pattern depends on
- Ratio between height and width of the bed
- Bed material size and density
- Gas velocity
- Air distributor

The contribution from circulation is usually embedded in D_{sr}

Splash zone is neglected (low velocity)

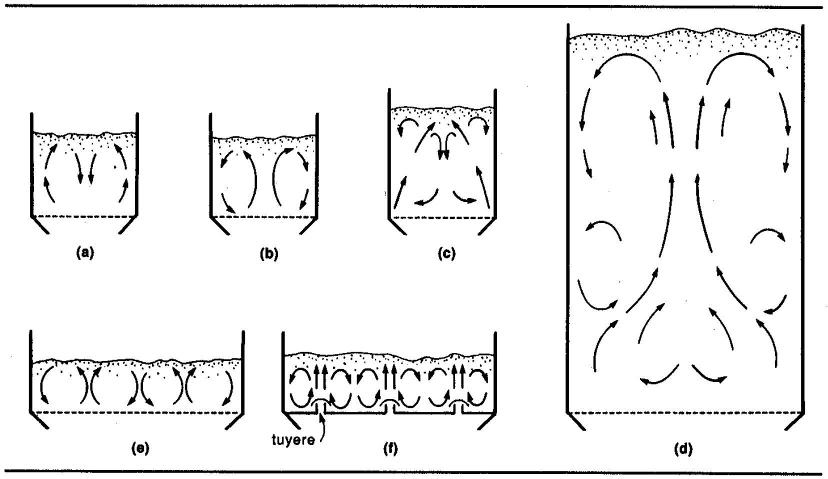


Figure from Kunii & Levenspiel, 1991, [2]

D_{sr} from wake-emulsion interchange

Kunii & Levenspiel, 1969 [1]

Theoretical deduction of the dispersion coefficient

- The rate of solids entering the wake (via cloud) of bubbles
- The mean square of lateral displacement by the wake

$$D_{sr} = \frac{3}{16}\left(\frac{\delta}{1-\delta}\right)\frac{U_{mf}d_b}{\varepsilon_{mf}}$$

δ - bubble fraction
ε - voidage
d_b -bubble diameter
U_{mf} - min. fluid. vel.

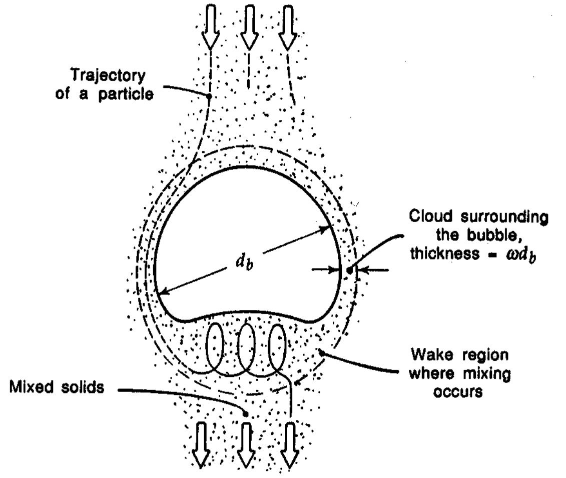


Figure from Kunii & Levenspiel, 1991 [2]

- Verified by two reported experiments
- Bubble diameter as fitting parameter
- Does not account for splash zone

D_{sr} in range of 0.0001 to 0.006 m^2/s

Bubbles

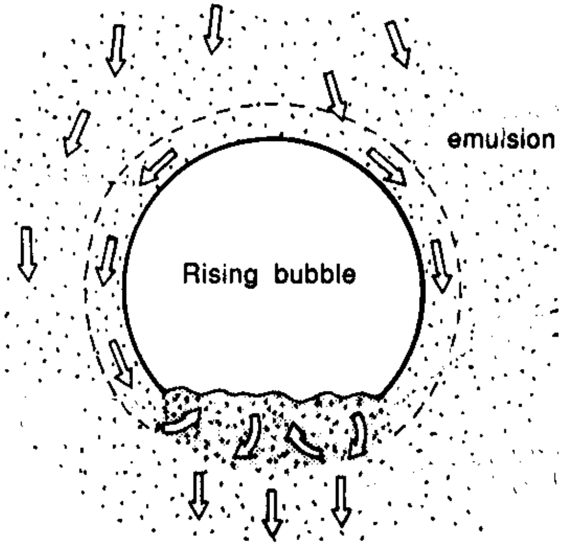


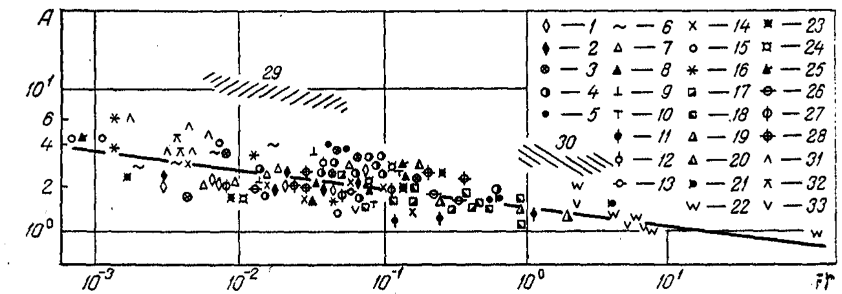
Figure from Kunii & Levenspiel, 1991 [2]

D_{sr} from heated tracer particles

Borodulya et al., 1982 [3] summarized an extensive series of experiments by the expression

$$\frac{D_{sr}}{\left(U - U_{mf}\right) H_0} = k_2 \left(\frac{D_c}{H_0}\right)^{n_1} Fr^{n_2}$$

k_2=0.013; n_1=0.5; n_2= -0.15



Experiments by heated tracer particles (baffled)
Wide range of particle sizes
Four different reactors

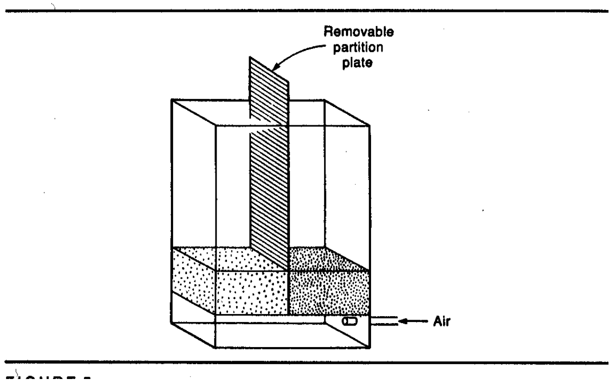


Figure from Kunii & Levenspiel, 1991

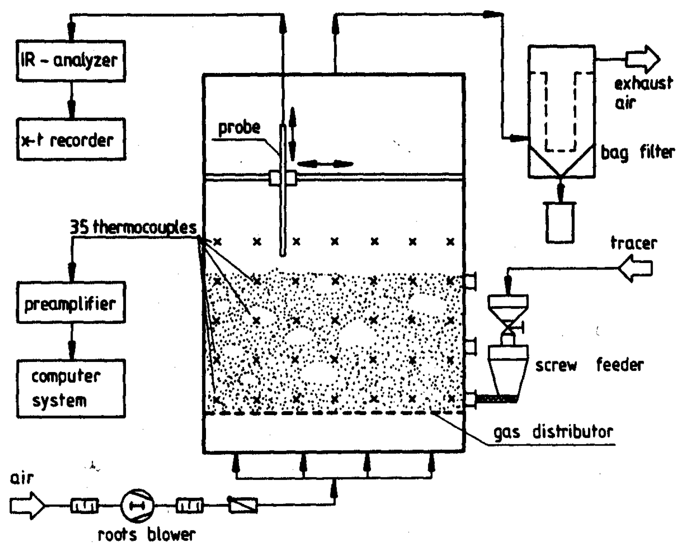
W = 0.4 – 0.7 m
H = 0.05 – 0.4 m
d_p = 0.25-1.75 mm
U-U_{mf} = 0.03–1.2 m/s

U – Gas velocity
H_0 – Bed height
D_c – Diameter of vessel
Fr – Froude number $(u\text{-}u_0)^2/gH_0$

D_{sr} from frozen CO_2 as tracer particles

Bellgardt & Werther, 1985 [4], injected frozen CO_2
Estimated D_{sr} from CO_2 measurements over the bed

$$D_{sr} = D_0 + 0.023 \frac{1}{H} \int_0^H \frac{\delta}{1-\delta} \sqrt{g d_b^3}\, dh$$



D_0=0.67e-3 m²/s

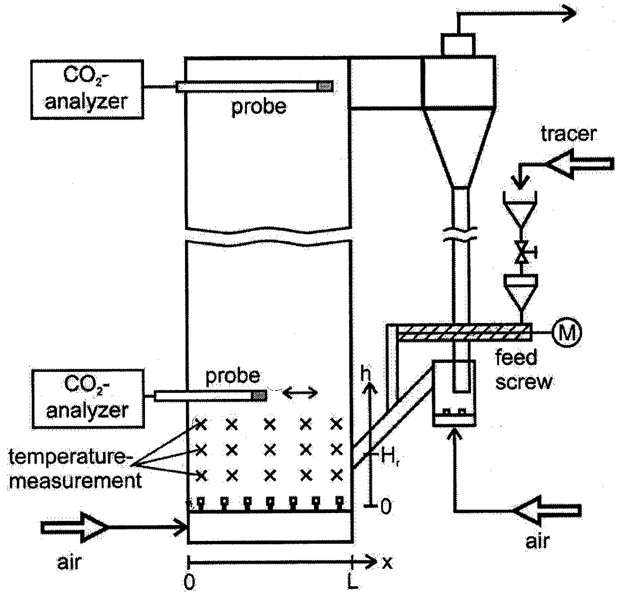
δ - bubble fraction
d_b -bubble diameter
H – bed height

W = 2 m
H = 0.4 - 0.9 m
d_p = 0.45, 0.85 mm
U-U_{mf} = 0.1–0.7 m/s

D_{sr} in range of 0.0007 to 0.003 m^2/s

D_{sr} in CFB (frozen CO_2)

Sclichthaerle & Werther, 2001 [5]
- Same technique as in Bellgardt et al.
- Under CFB conditions
- Added lateral convective flow
- $D_{sr} \approx 0.12$ m^2/s

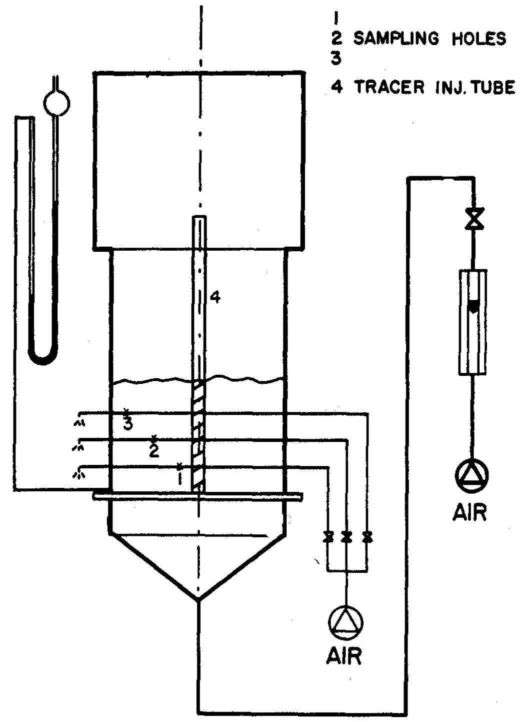


$W = 1.0$ m
$H = 0.8$ m
$d_p = 0.15$ mm
$U\text{-}U_{mf} = 3$ m/s

D_{sr} from sampling of salt as tracer particles

Berruti et al., 1986 [6], measured tracer particle concentrations.

- Cylindrical reactor
- Cylindrical baffle
- Three different heights
- Various distances from the centre
- D_{sr} varies with location in the bed



$$D_{sr} = 0.185 \cdot 10^{-4} \cdot \left[1 - \left(0.44 + 2.87\frac{h}{H_{mf}}\right)\left(\frac{r}{R}\right)^5\right] \cdot (U - U_{mf})d_p$$

$$\cdot \left[\frac{(U - U_{mf})d_p \rho_f}{\mu_f}\right]^{-0.25} \cdot \left[\frac{h}{d_p}\right]^{1.45} \cdot \left[\frac{\rho_p - \rho_f}{\rho_f}\right]^{-0.43}$$

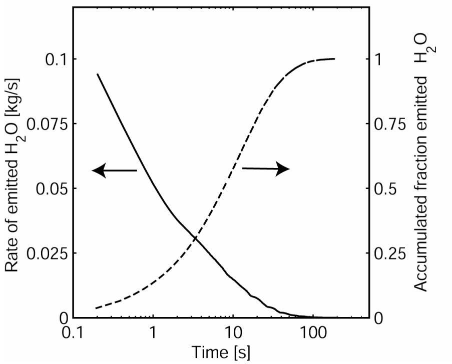
U – Gas velocity
H_{mf} – Bed height at min. fluid. Vel.
R – Radius of vessel
d_p – particle diameter
ρ_f – density of fluid
ρ_p – density of solids
μ_f – viscosity of fluid

$W = 0.27$ m
$H = 0.19$ m
$d_p = 0.40$ mm
$U\text{-}U_{mf} = 0.1$–0.2 m/s

D_{sr} in range of 0.0002 to 0.002 m^2/s

D_{sr} from modelled drying rate of fuel and measured concentrations of H_2O above the bed [7]

Drying rate + Dispersion of fuel → Distribution of H_2O above the bed



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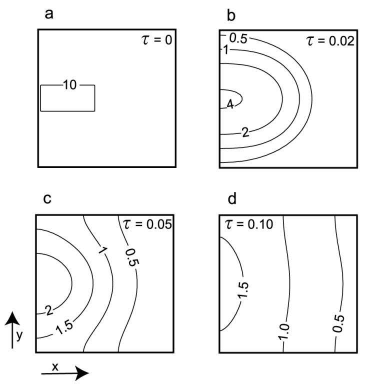

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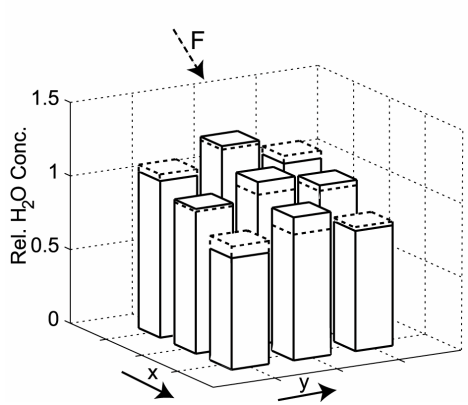
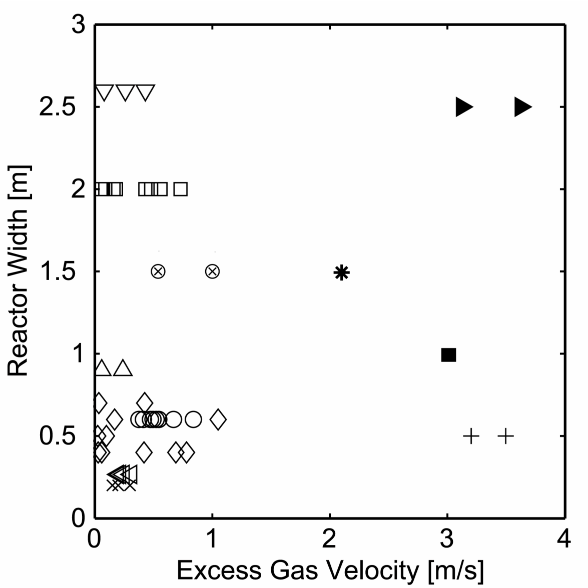
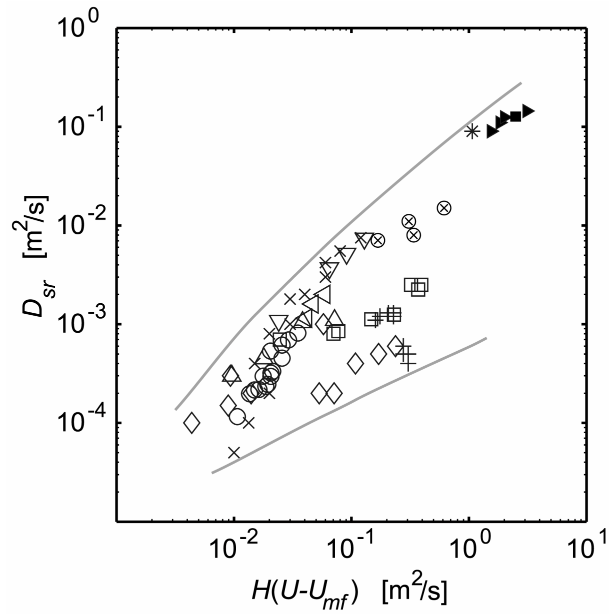


Comparison between modelled and measured concentrations of H_2O gives best fit for Dsr ≈ 0.1 m^2/s

$W = 1.5$ m
$H = 0.5$ m
$d_p = 0.7$ mm
$U\text{-}U_{mf} = 2.1$ m/s

Experimental conditions and dispersion coefficient – literature data





Test conditions and results, filled symbols for circulating fluidised beds. (⊗) Borodulya et al. [3], (③), Bellgardt and Werther [4], (■) Schlichthaerle and Werther [5] (⅄) Berutti et al. [6], (*) Niklasson et al. [7], (+) Bi et al. [8], (□) Highley and Merrick [9], (Δ) Salam et al. [10], (◇) Subbarao et al. [11], (O) Yan et al. [12], (∇) Xiang et al. [13] and (►) Xiao et al. [14].

Conclusion

- D_{sr} varies between 0.0001 and 0.1 m^2/s
- Empirical formulas from experiments in lab-scale reactors are not applicable to boiler conditions
- D_{sr} depends on the location in the bed

Future work

- Experiments to study D_{sr} in the splash zone
- Establish a 3D model, which includes convection

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Counter Current Back Mixing

van Deemter

Model of vertical mixing of solids

$$\begin{cases} f_d \dfrac{\partial C_{sd}}{\partial t} + f_d u_{sd} \dfrac{\partial C_{sd}}{\partial z} + K_s \left(C_{sd} - C_{su} \right) = 0 \\ f_u \dfrac{\partial C_{su}}{\partial t} + f_u u_{su} \dfrac{\partial C_{su}}{\partial z} + K_s \left(C_{su} - C_{sd} \right) = 0 \end{cases}$$

$$K_s = \frac{\text{Volume transfer rate from emulsion to wake}}{\text{Volume of bubble}}$$

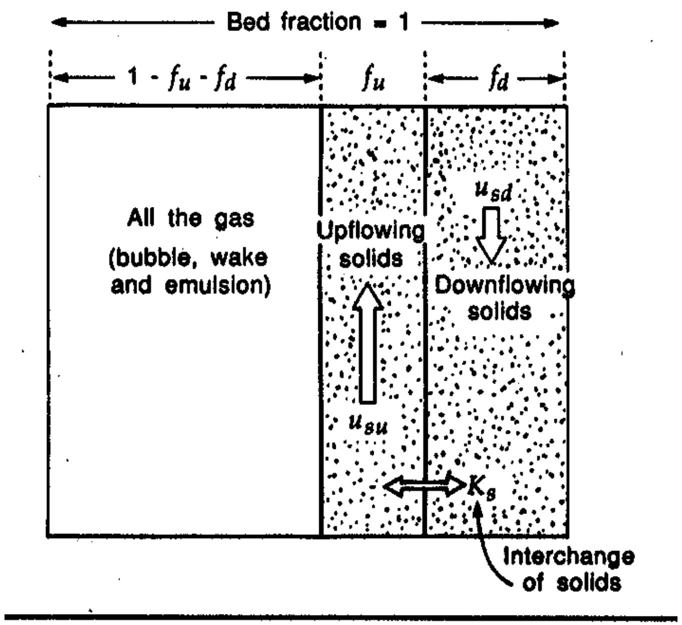

Figure from Kunii & Levenspiel, 1991