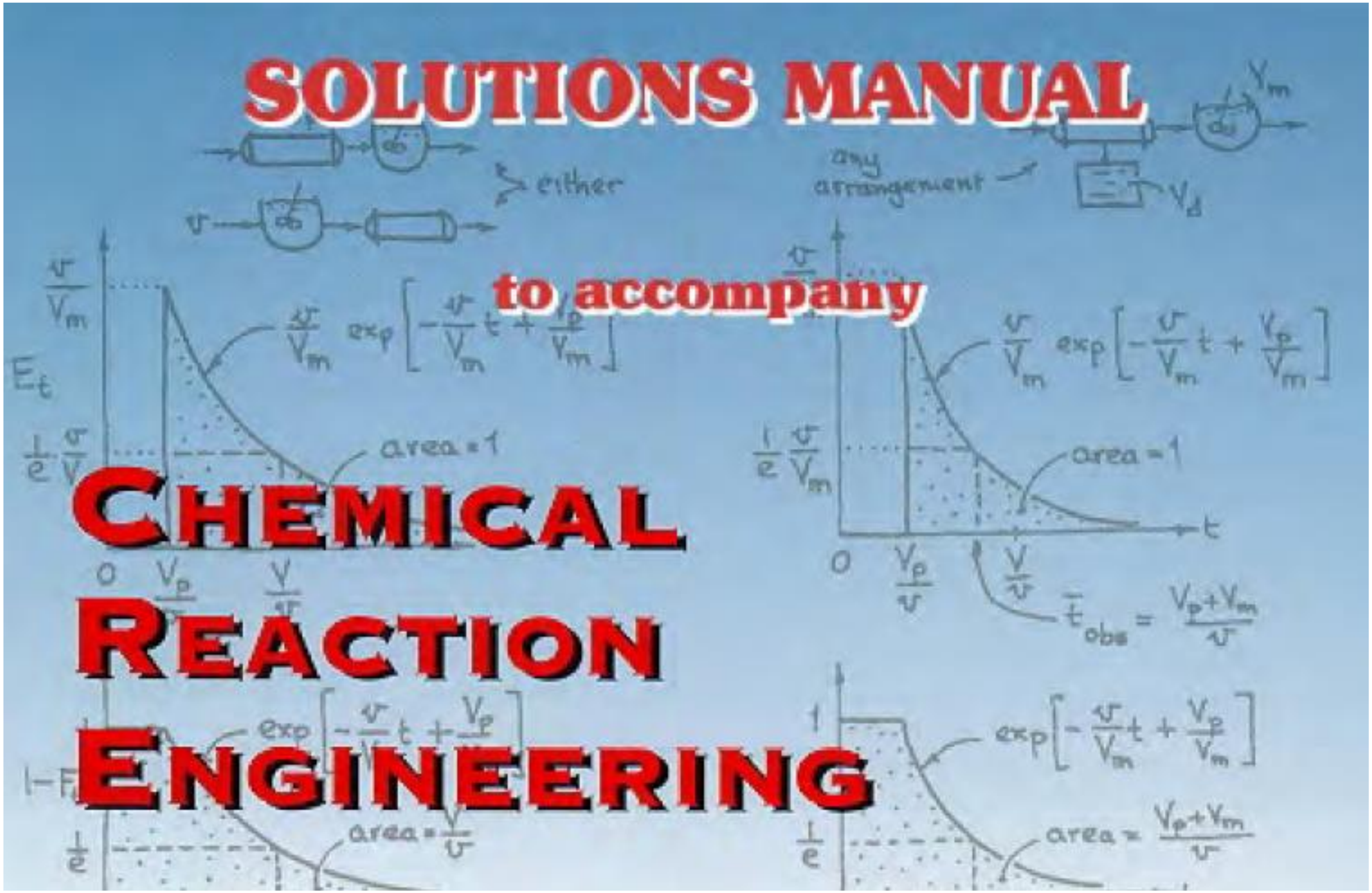
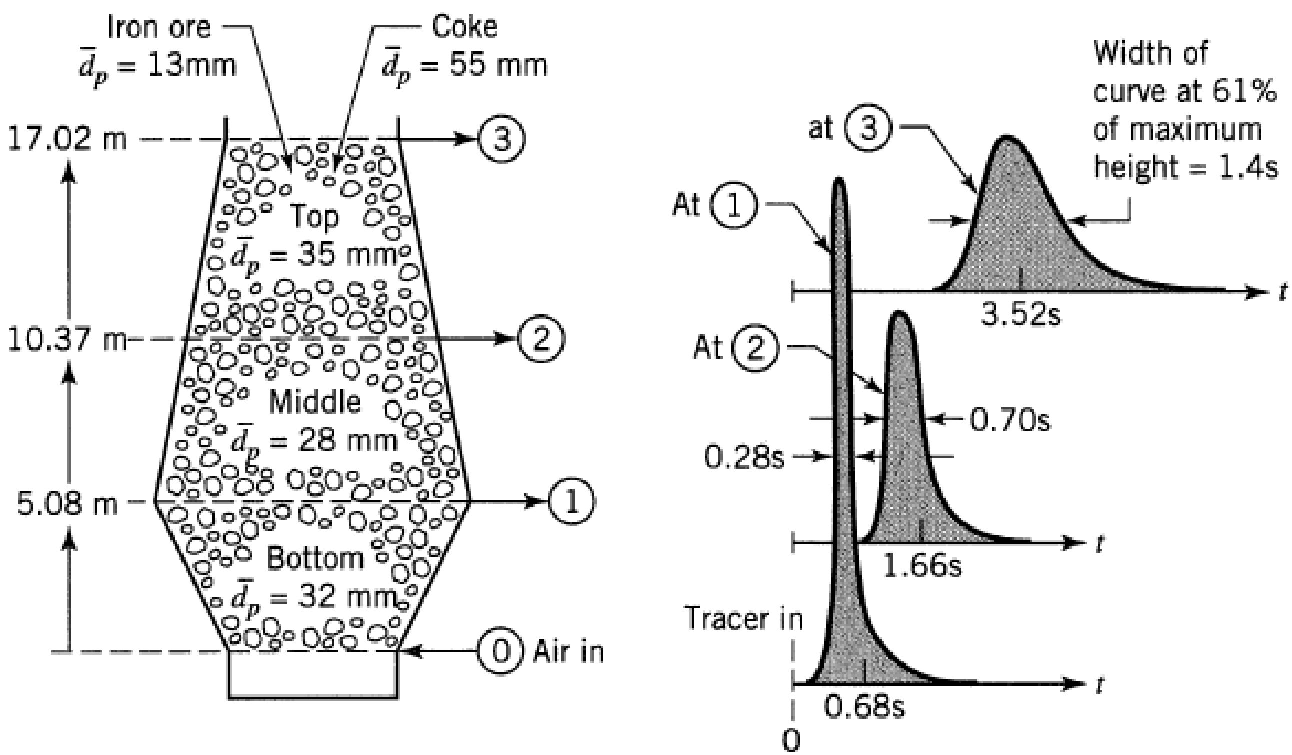
حل چند مساله

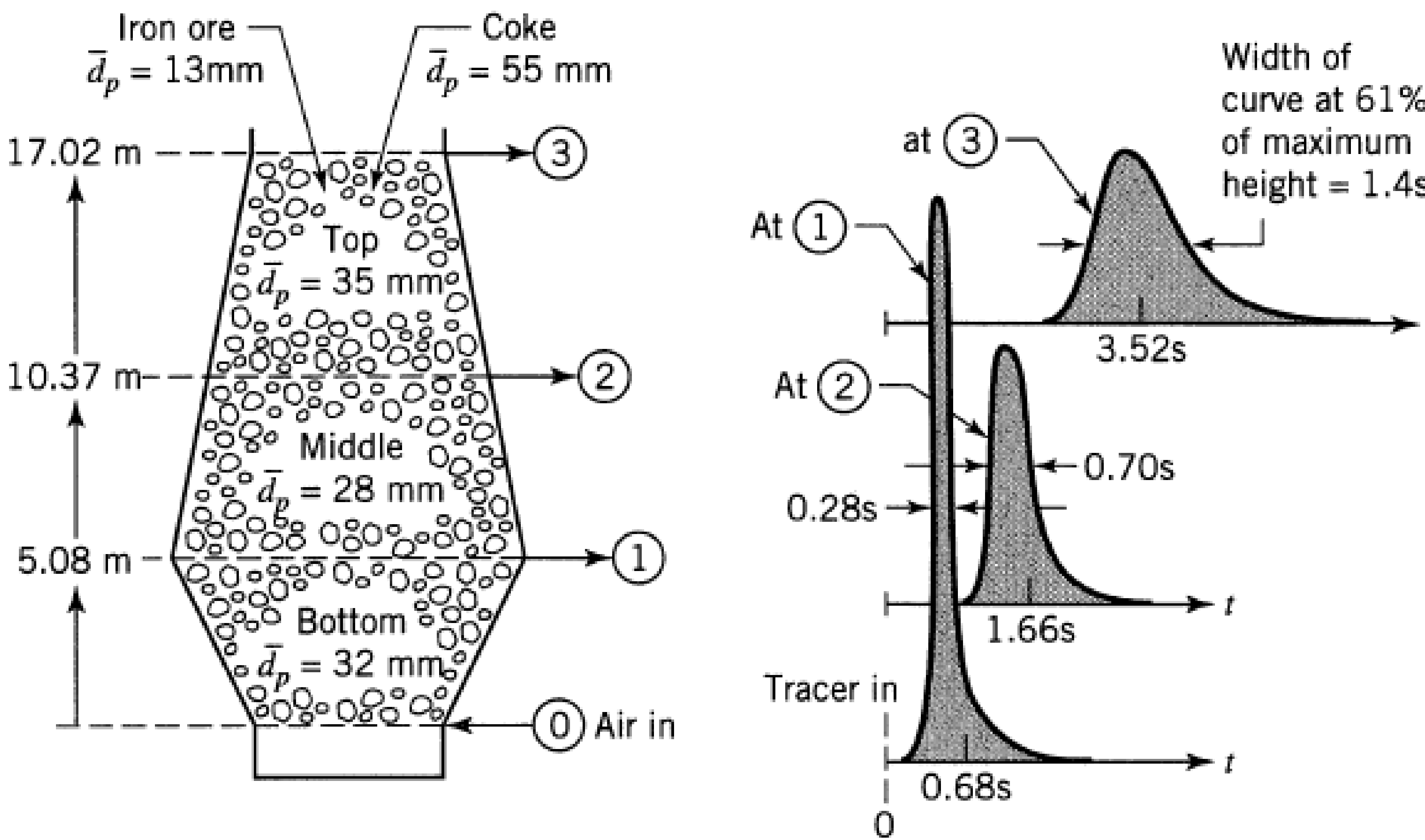

SOLUTIONS MANUAL
to accompany
CHEMICAL
REACTION
ENGINEERING
either
any arrangement
area = 1
area = 1

13.1. The flow pattern of gas through blast furnaces was studied by VDEh (Veren Deutscher Eisenhüttenleute Betriebsforschungsinstitut) by injecting Kr-85 into the air stream entering the tuyeres of the 688 m³ furnace. A sketch and listing of pertinent quantities for run 10.5 of 9.12.1969 is shown in Fig. P13.1. Assuming that the axial dispersion model applies to the flow of gas



in the blast furnace, compare **D**/*ud* for the middle section of the blast furnace with that expected in an ordinary packed bed.

From Standish and Polthier, *Blast Furnace Aerodynamics,* p. 99, N. Standish, ed., Australian I. M. M. Symp., Wollongong, 1975.



Iron ore
$\bar{d}_p$ = 13mm
Coke
$\bar{d}_p$ = 55 mm
17.02 m
10.37 m
5.08 m
3
2
1
0 Air in
Top
$\bar{d}_p$ = 35 mm
Middle
$\bar{d}_p$ = 28 mm
Bottom
$\bar{d}_p$ = 32 mm
Width of curve at 61% of maximum height = 1.4s
at 3
At 1
At 2
3.52s
0.70s
0.28s
1.66s
Tracer in
0.68s
0
t

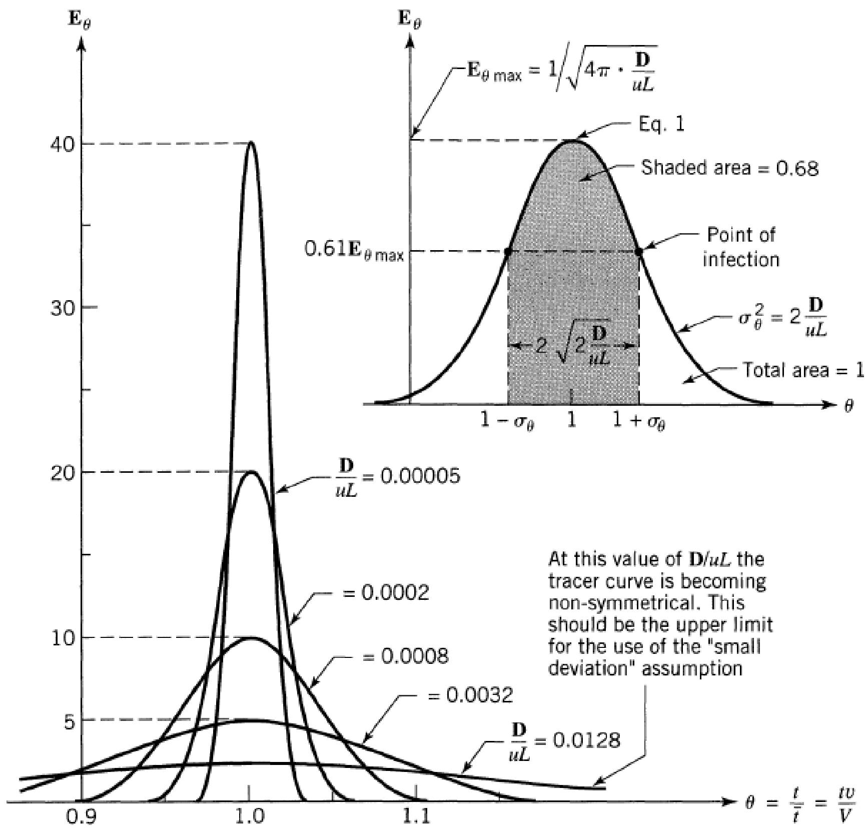
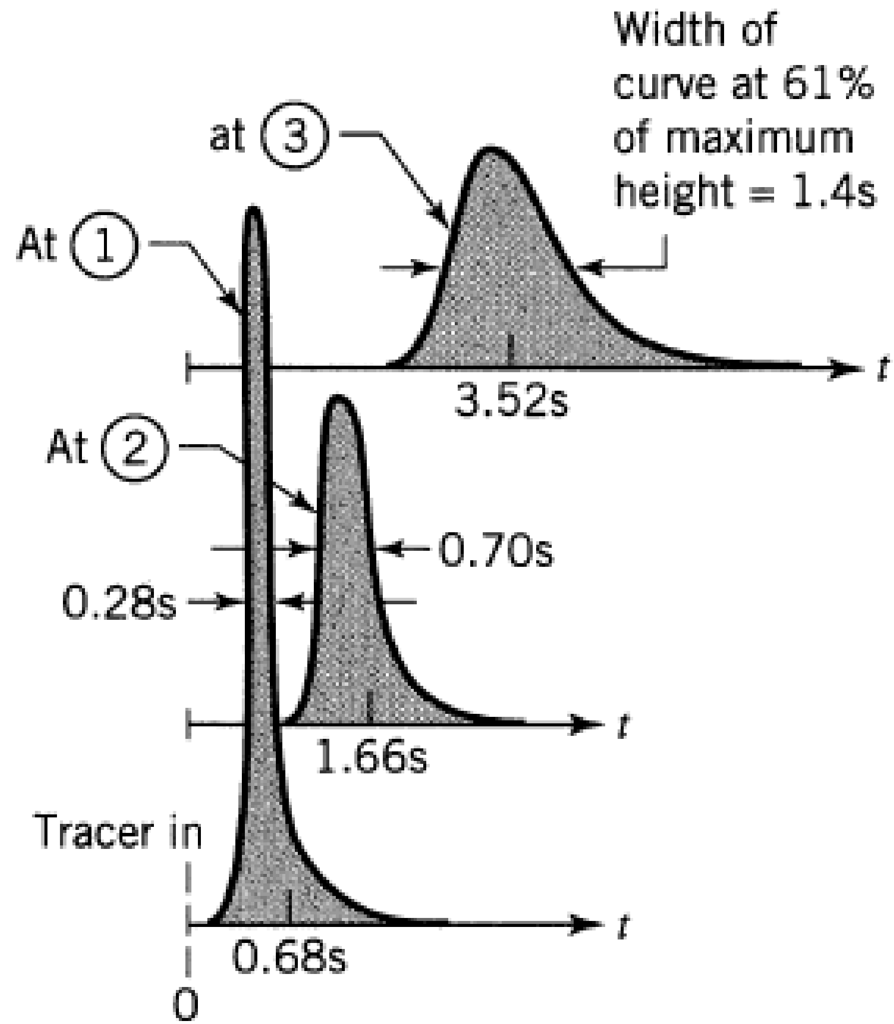


Figure 13.4 Relationship between $\mathbf{D}/uL$ and the dimensionless $\mathbf{E}_\theta$ curve for small extents of dispersion, Eq. 7.



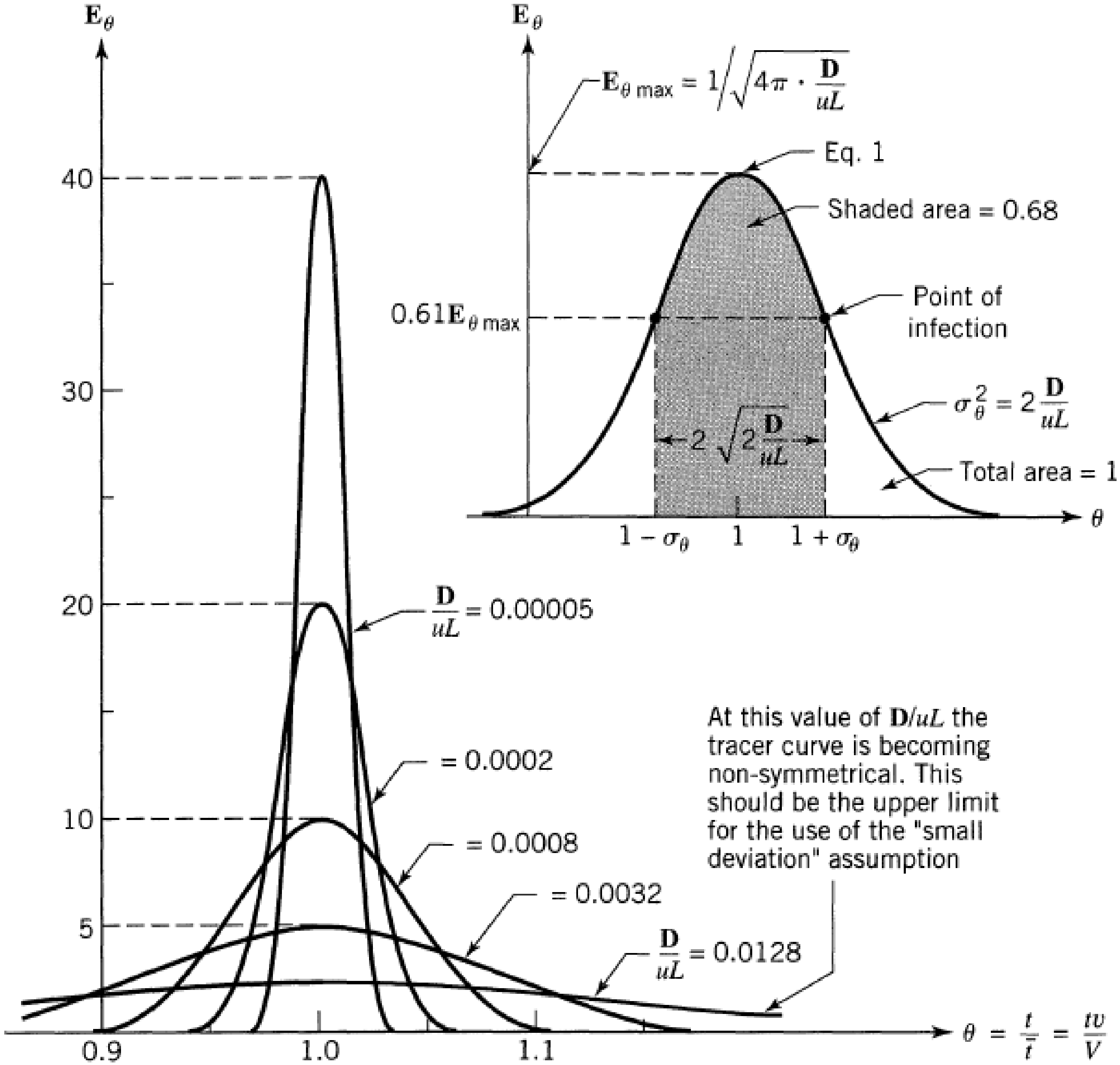


Figure 13.4 Relationship between $\mathbf{D}/uL$ and the dimensionless $\mathbf{E}_\theta$ curve for small extents of dispersion, Eq. 7.

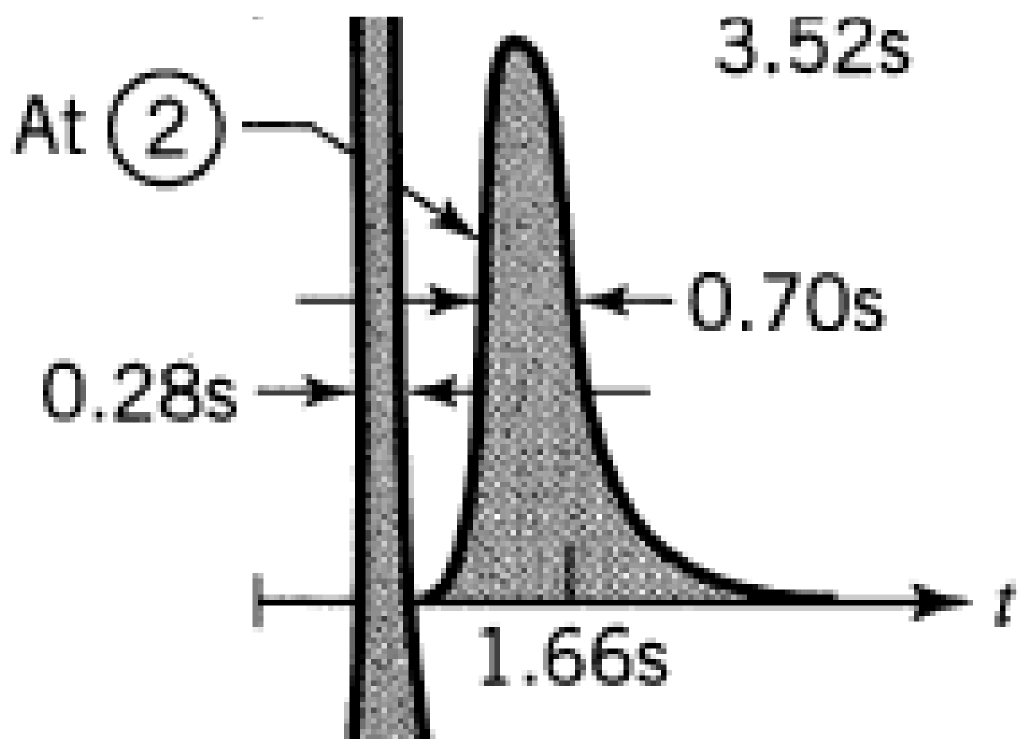
13.1 Evaluate the intensity of dispersion for the middle section.

$$\bar{t}_2 = 1.66\,s$$

Width at 61% = 0.70

$\therefore$ standard deviation, $\sigma_2 = 0.35$ (from Fig 4)

$$\sigma_2^2 = 0.1225$$



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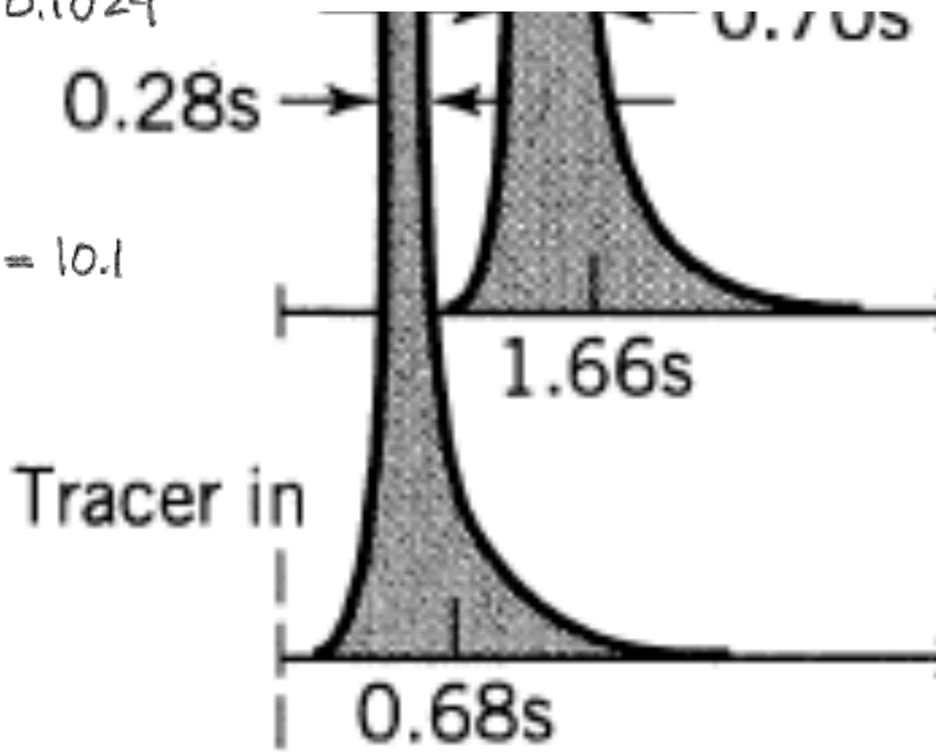
$$\Delta\bar{t} = 1.66 - 0.068 = 0.98 \text{ (from Fig P1)}$$

$$\Delta(\sigma^2) = 0.1225 - 0.0196 = 0.1029$$

$$\therefore \frac{D}{uL} = \frac{\Delta(\sigma^2)}{\Delta(\bar{t})^2} = \frac{0.1029}{2(0.98)^2} = 0.0536$$

$$\frac{D}{ud_p} = \frac{D}{uL}\left(\frac{L}{d_p}\right) = (0.0536)\left(\frac{10.37 - 5.08}{0.028}\right) = 10.1$$



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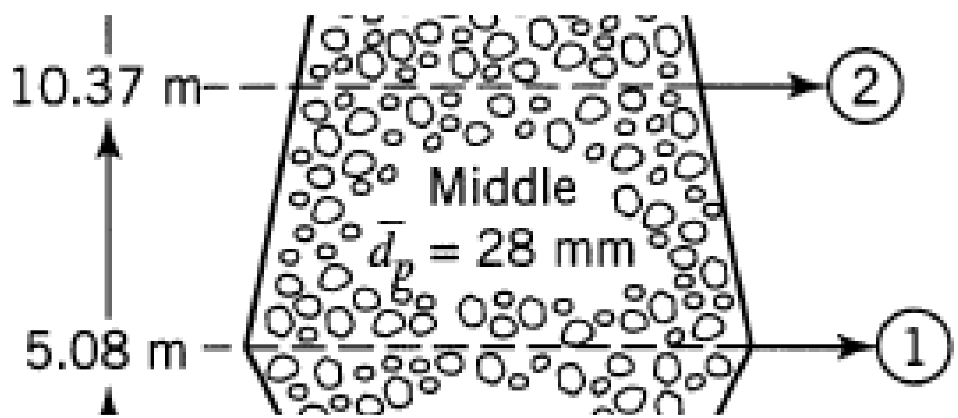
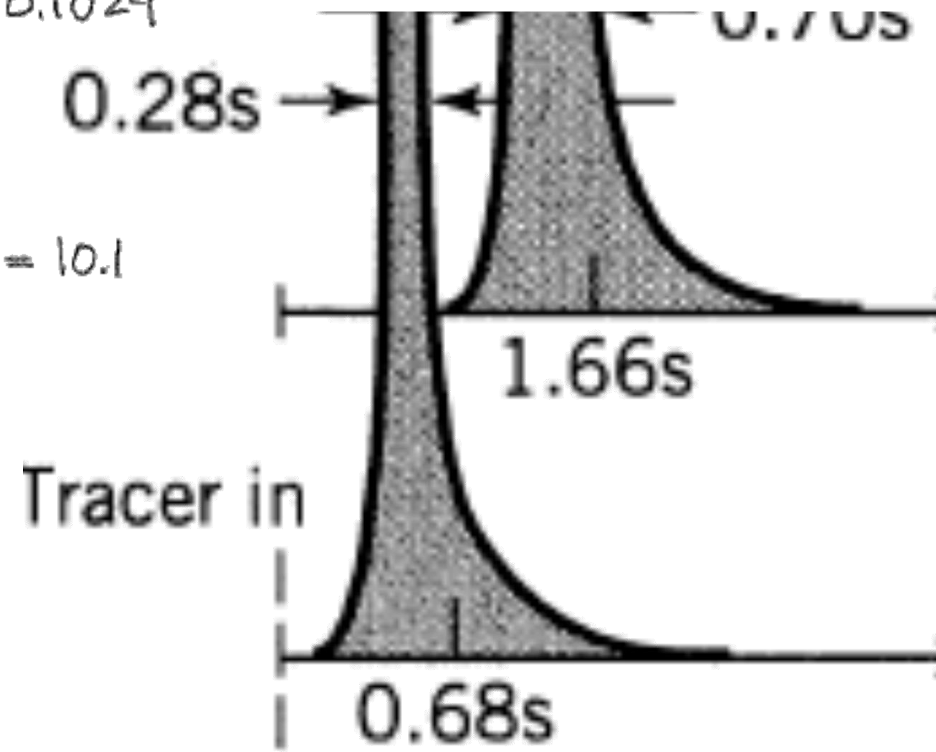




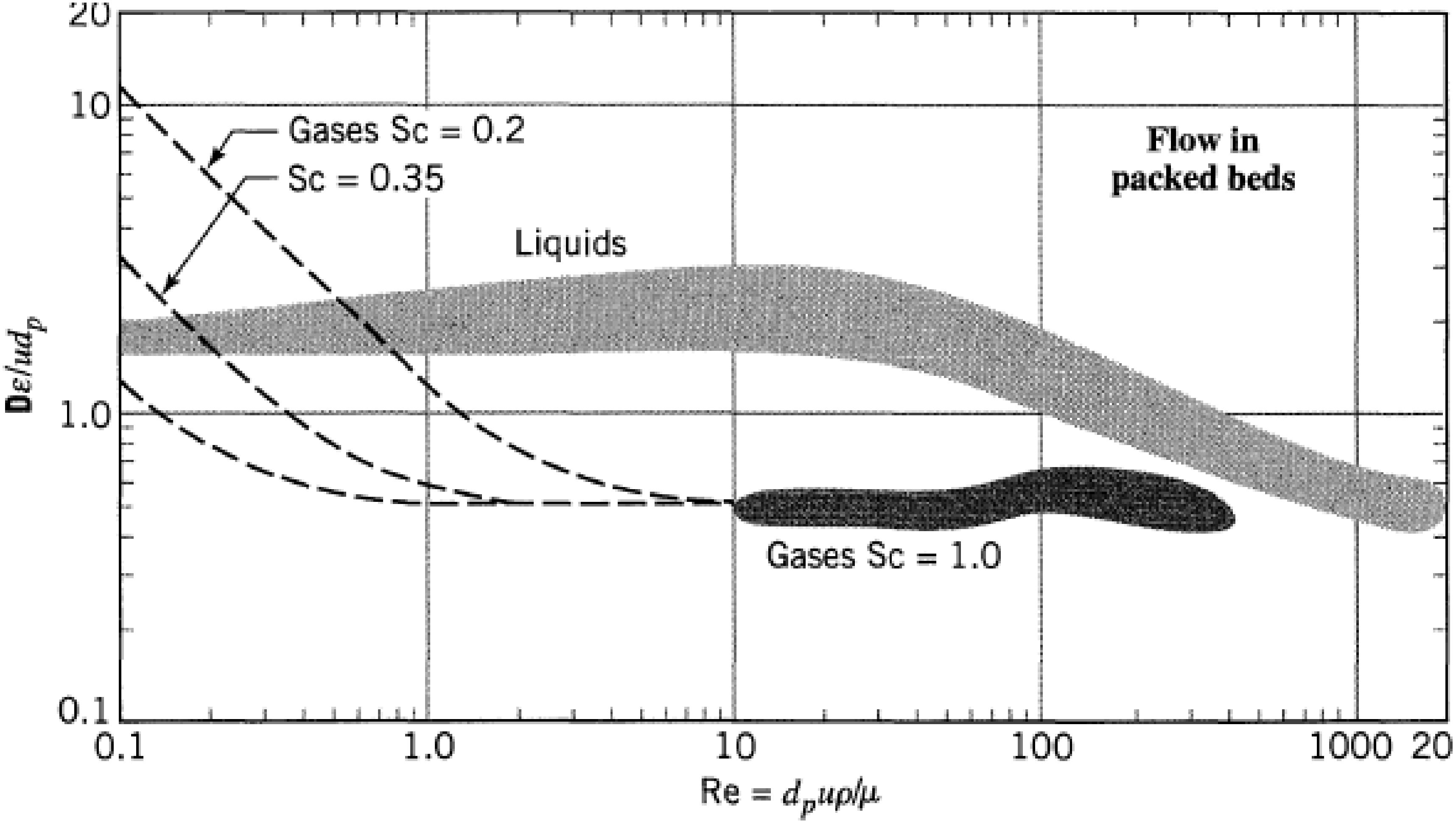
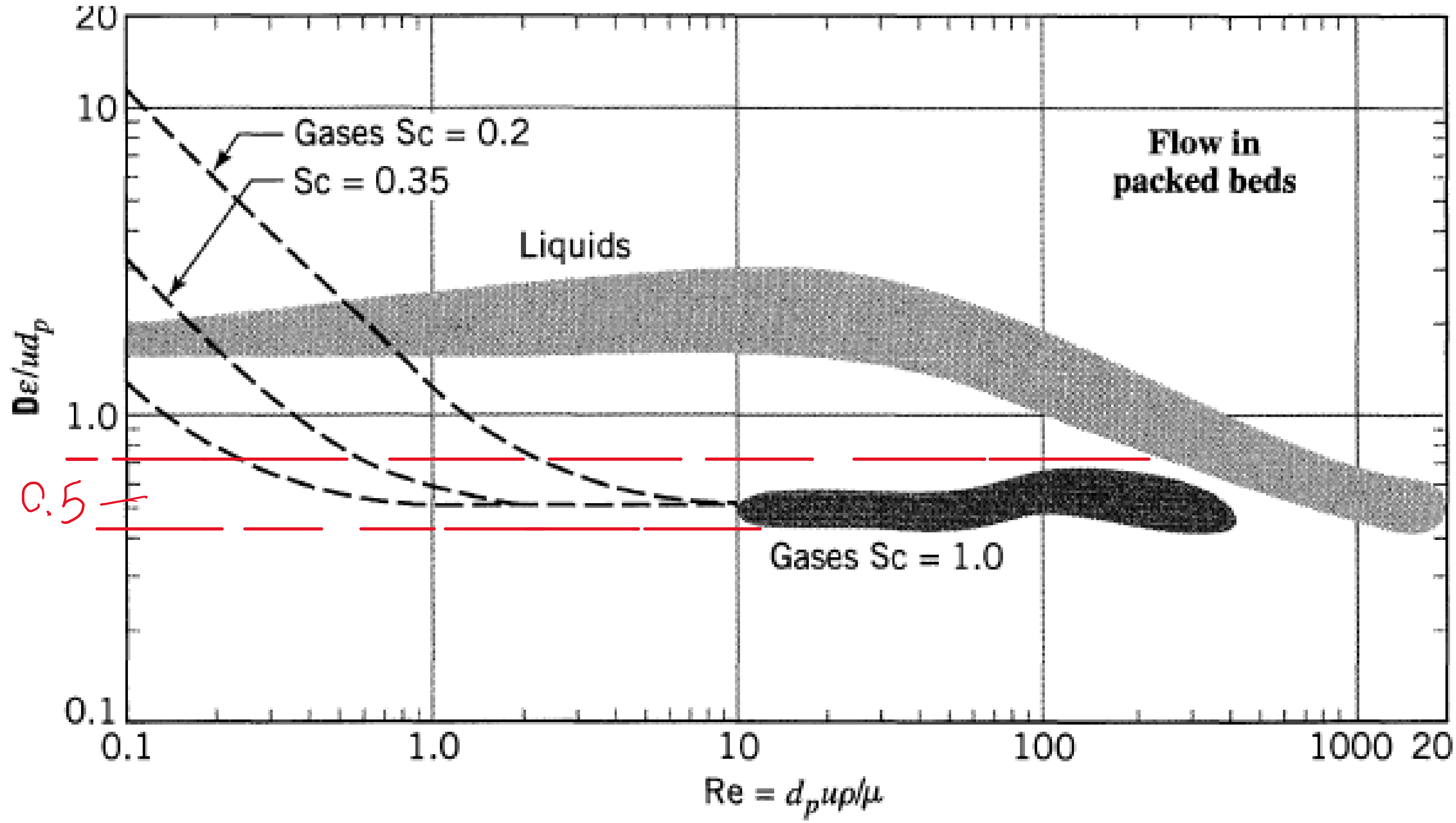


Figure 13.17 Experimental findings on dispersion of fluids flowing with mean ax velocity u in packed beds; prepared in part from Bischoff (1961).



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$$\frac{D}{ud_p} = \frac{D}{uL}\left(\frac{L}{d_p}\right) = (0.0536)\left(\frac{10.37 - 5.08}{0.028}\right) = 10.1$$

From the blast furnace experiment $\frac{D}{ud_p} = 10.1$

From the correlation for smaller solids, Fig 17 : $\frac{D}{ud_p} = 0.5$

Note: These results are very different. It could be because of the severe bypassing in the blast furnace, caused by the segregation of the solids and the severe channeling of the gas

13.2. Denmark's longest and greatest river, the Gudenaa, certainly deserves study, so pulse tracer tests were run on various stretches of the river using radioactive Br-82. Find the axial dispersion coefficient in the upper stretch of the river, between Tørring and Udlum, 8.7 km apart, from the following reported measurements.

t, hr	C, arbitrary	t, hr	C, arbitrary
3.5	0	5.75	440
3.75	3	6	250
4	25	6.25	122
4.25	102	6.5	51
4.5	281	6.75	20
4.75	535	7	9
5	740	7.25	3
5.25	780	7.5	0
5.5	650		

Data from Danish Isotope Center, report of November 1976.

حل : جدول را کامل کنیم

t_i	C_i	$t_i C_i$	$t_i^2 C_i$	...

$$\bar{t} = \frac{\int_0^\infty C\,t\,dt}{\int C\,dt} = \frac{\sum C_i t_i}{\sum C_i} = 5.25\,hr$$

$$\sigma^2 = \frac{\sum C_i t_i^2 \Delta t}{\sum C_i \Delta t} - \bar{t}^2 = \cdots = 0.282\,(hr)^2$$

$$\sigma^2_{(\theta)} = \frac{\sigma^2(t)}{\bar{t}^2} = 0.0102$$

If $\frac{D}{uL} < 0.01 \rightarrow \sigma^2(\theta) = \frac{2D}{uL}$

$$\hookrightarrow \frac{2D}{uL} = 0.0102$$

$$\hookrightarrow \frac{D}{uL} = 0.0051 < 0.01 \quad \text{OK}$$

شرایط ظرف بسته وجود دارد

$L = 8.7\ km$ فاصله دو شهر

$\frac{D}{u(8.7)} = 0.0051 \rightarrow \frac{D}{u} = 0.0444\ km$

$u = \frac{L}{t} = \frac{8.7\ km}{5.25\ hr}$ سرعت خطی رود آب

$= 1.657\ \frac{km}{hr}$, $\boxed{D = 0.07357}$

13.3. RTD studies were carried out by Jagadeesh and Satyanarayana (IEC/PDD **11** 520, 1972) in a tubular reactor (L = 1.21 m, 35 mm ID). A squirt of NaCl solution (5 N) was rapidly injected at the reactor entrance, and mixing cup measurements were taken at the exit. From the following results calculate the vessel dispersion number; also the fraction of reactor volume taken up by the baffles.

t, sec	NaCl in sample
0–20	0
20–25	60
25–30	210
30–35	170
35–40	75
40–45	35
45–50	10
50–55	5
55–70	0

(v = 1300 ml/min)

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1

2

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3

13.3 From experiment:

t, sec	$\bar{t}$, sec	C
0-20	10	0
20-25	22.5	60
25-30	27.5	210
30-35	32.5	170
35-40	37.5	75
40-45	42.5	35
45-50	47.5	10
50-55	52.5	5
55-70	62.5	0

13.3 From experiment:

t, sec	$\bar{t}$, sec	C
0-20	10	0
20-25	22.5	60
25-30	27.5	210
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35-40	37.5	75
40-45	42.5	35
45-50	47.5	10
50-55	52.5	5
55-70	62.5	0

Calculate D/uL:

$\Sigma c = 565$

$\Sigma tc = 17687.5$

$\Sigma t^2 c = 573781.25$

$$\bar{t} = \frac{\Sigma tc}{\Sigma c} = 31.31 \text{ sec}$$

$$\sigma^2 = \frac{\Sigma t^2 c}{\Sigma c} - (\bar{t})^2 = 35.52 \text{ sec}^2$$

$$\frac{D}{uL} = \frac{\sigma^2}{2(\bar{t})^2} = 0.018 \longleftarrow$$

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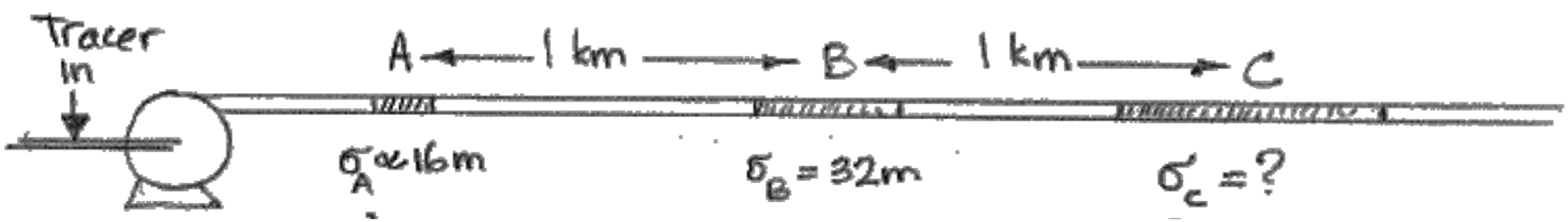
Now calculate the volume of the baffles

From mat. balance $\left.\begin{array}{l} V = (1.21\times10^2)(\pi \times 1.75^2) = 1164 \text{ cm}^3 \\ v = 1300 \text{ cm}^3/\text{min} = 21.67 \text{ cm}^3/\text{s} \end{array}\right\} \bar{t} = \dfrac{1164}{21.67} = 53.72 \text{ sec}$

$\therefore$ Volume of baffles $= \dfrac{53.72 - 31.31}{53.72}(1164) = 486 \text{ cm}^3$

13.5. An injected slug of tracer material flows with its carrier fluid down a long, straight pipe in dispersed plug flow. At point *A* in the pipe the spread of tracer is 16 m. At point *B*, 1 kilometer downstream from A, its spread is 32 m. What do you estimate its spread to be at a point *C*, which is 2 kilometers downstream from point A?



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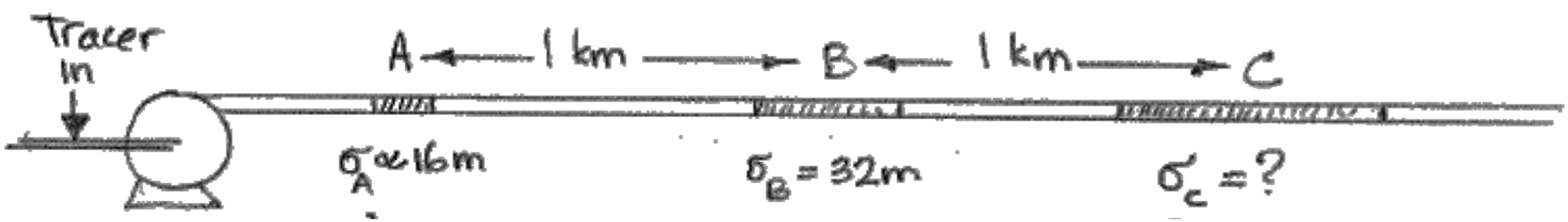
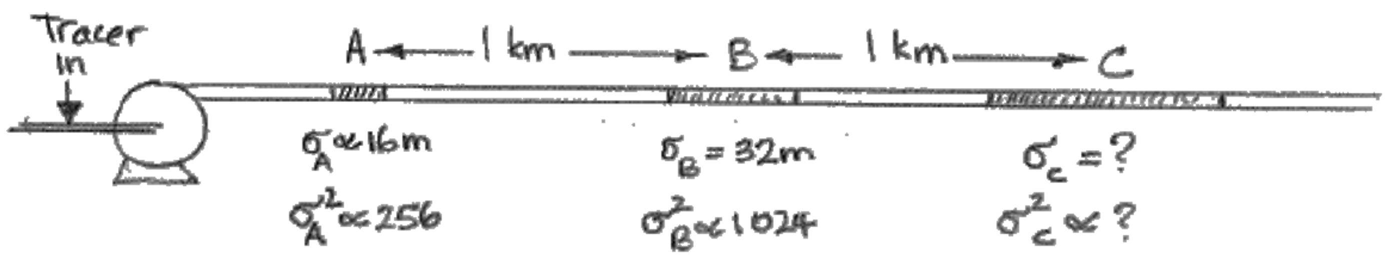


$$\sigma^2 \propto L \qquad \text{or} \qquad \left(\begin{matrix}\text{width of tracer}\\ \text{curve}\end{matrix}\right)^2 \propto L$$

13.5



Let us solve this with the dispersion model. We could also do this with the tanks-in-series model.

Then from Eq 8 $\quad \sigma^2 = 2\left(DL/u^3\right)$

$$\text{or} \quad \sigma^2 \propto L \quad \cdots \quad \text{or } (\text{spread})^2 \propto L$$

So $\quad \sigma_c^2 - \sigma_B^2 = \sigma_B^2 - \sigma_A^2$

or $\quad \sigma_c^2 - 1024 = 1024 - 256$

or $\sigma_c^2 = 1792 \qquad\qquad$ or Width $= k\,\sigma_c = 42.3\text{m}$

13.10. A 12-m length of pipe is packed with 1 m of 2-mm material, 9 m of 1-cm material, and 2 m of 4-mm material. Estimate the variance in the output C curve for a pulse input into this packed bed if the fluid takes 2 min to flow through the bed. Assume a constant bed voidage and a constant intensity of dispersion given by $\mathbf{D}/ud_p = 2$.

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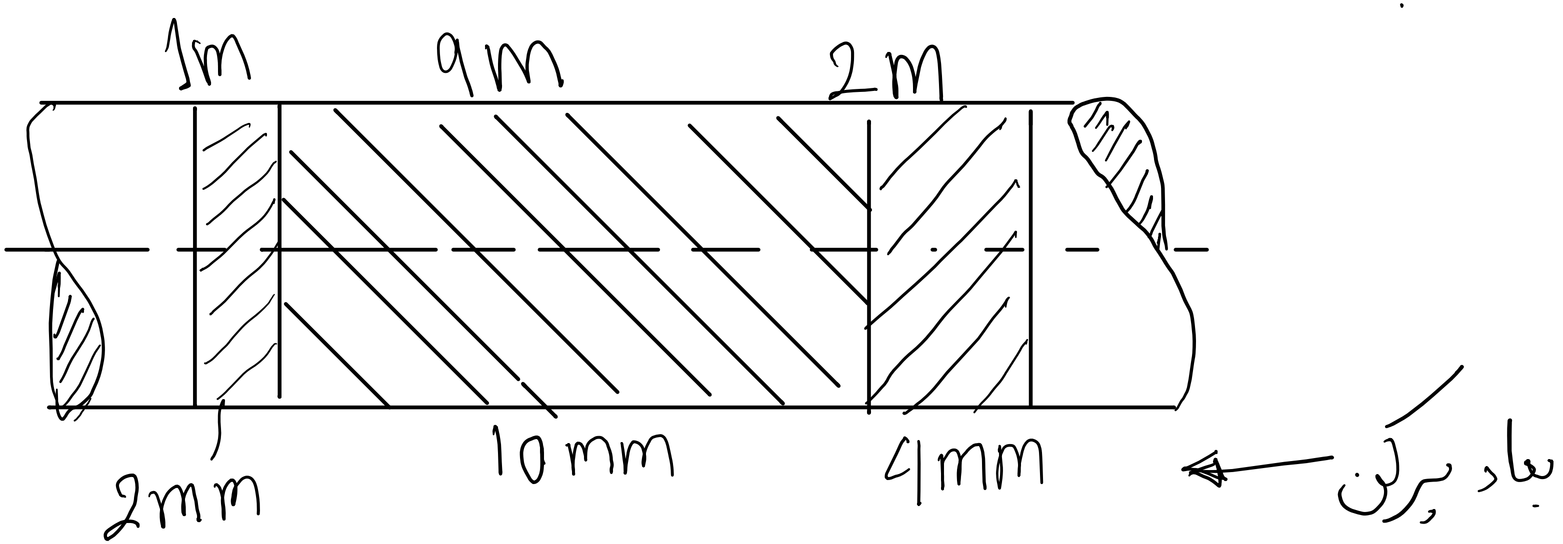

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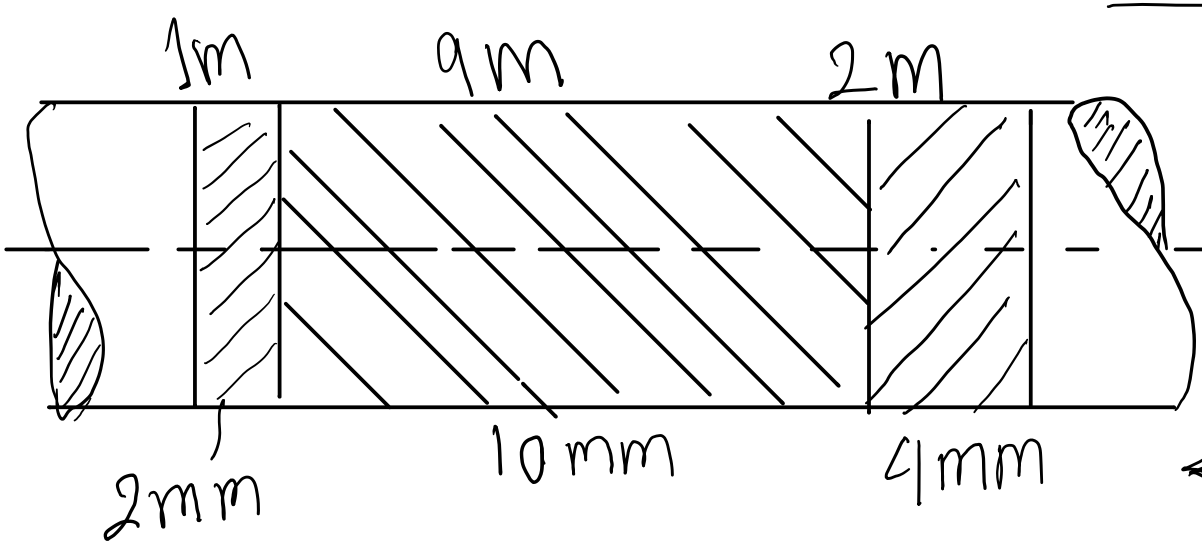
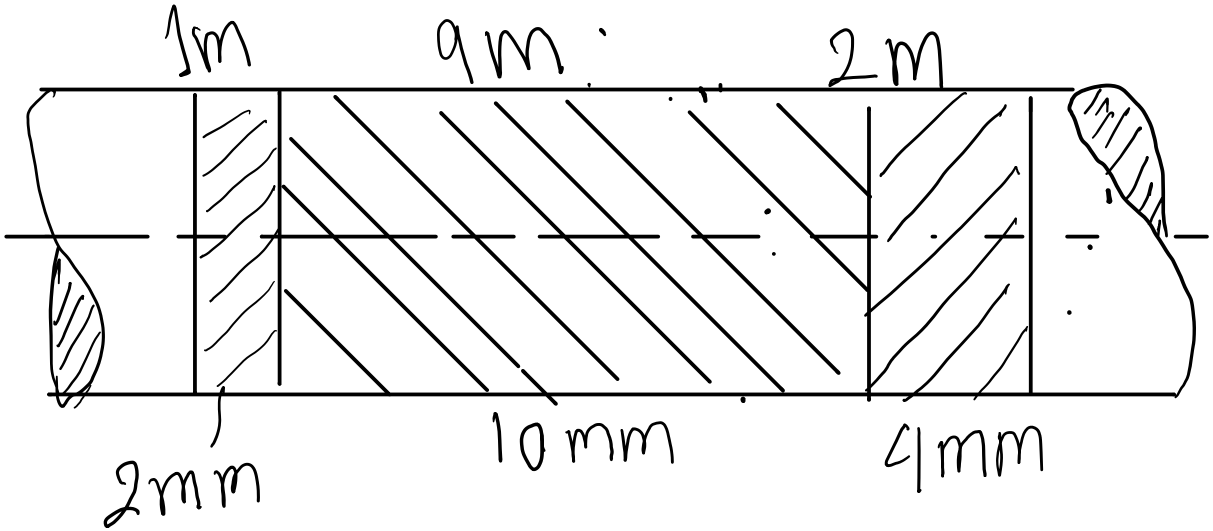
مجهول σ^2

با فرضیات فوق



ابعاد پرکن

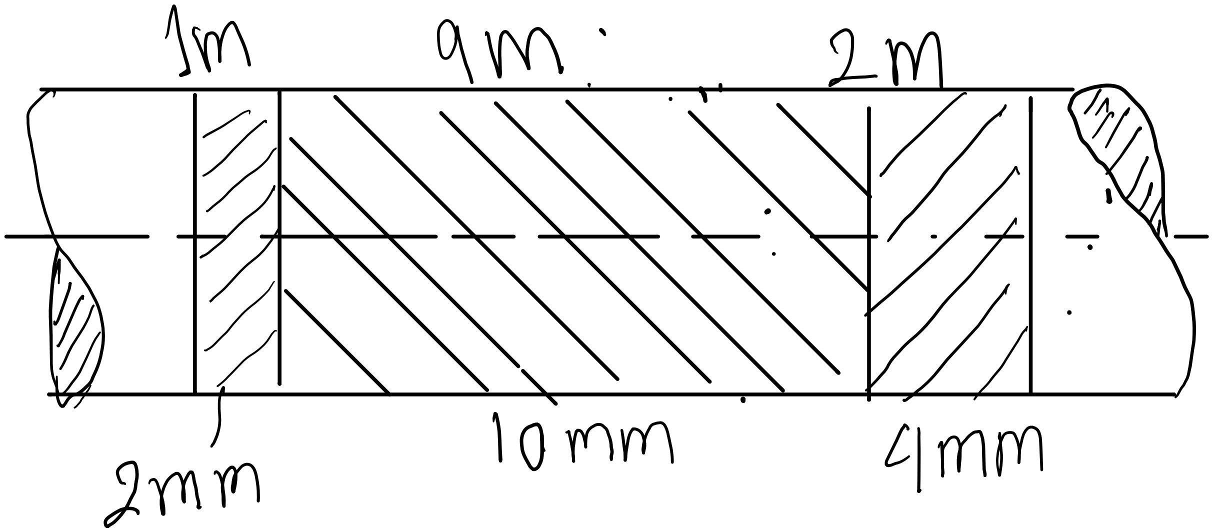
حل



$\bar{t} = 2\,min = 120\,sec$

$\frac{D}{u\,d_p} = 2$ ثابت

حل :



باتوجه به ثابت بودن فضای خالی در کل بستر

$L_1 = 1m \,, dp_1 = 2mm \,, t_1 = 10 sec$

$L_2 = 9m \,, dp_2 = 10mm \,, t_2 = 90 sec$

$L_3 = 2m \,, dp_3 = 4mm \,, t_3 = 20 sec$

$$\frac{D}{uL} = \left(\frac{D}{u d_p}\right)\left(\frac{d_p}{L}\right) = 2\left(\frac{d_p}{L}\right)$$

برای هر تست می‌توان $\frac{D}{uL}$ را بدست آورد.

تست ①: $2\left(\frac{2}{1000}\right) = 0.004 = \frac{D}{uL} < 0.01$

$$\sigma_1^2(\theta) = 2\left(\frac{D}{uL}\right)_1 = 0.008$$

زمان متوسط $\bar{t}_1 = 10\ \text{sec} \ \text{or}\ \frac{1}{6}\ \text{min}$

$$\sigma_1^2 = \bar{t}_1^2\, \sigma_1^2(\theta)$$

$$= \left(\frac{1}{6}\right)^2 (0.008) = 0.00022\ min^2$$

به همین ترتیب

$$\sigma_2^2 = 0.00999\ min^2$$

$$\sigma_3^2 = 0.00088\ min^2$$

$$\sigma_{کل}^2 = \sigma_1^2 + \sigma_2^2 + \sigma_3^2 = 0.011\ min^2$$

13.11. The kinetics of a homogeneous liquid reaction are studied in a flow reactor, and to approximate plug flow the 48-cm long reactor is packed with 5-mm nonporous pellets. If the conversion is 99% for a mean residence time of 1 sec, calculate the rate constant for the first-order reaction

(a) assuming that the liquid passes in plug flow through the reactor

(b) accounting for the deviation of the actual flow from plug flow

(c) What is the error in calculated k if deviation from plug flow is not considered?

Data: Bed voidage $\varepsilon = 0.4$

Particle Reynolds number $\mathrm{Re}_p = 200$

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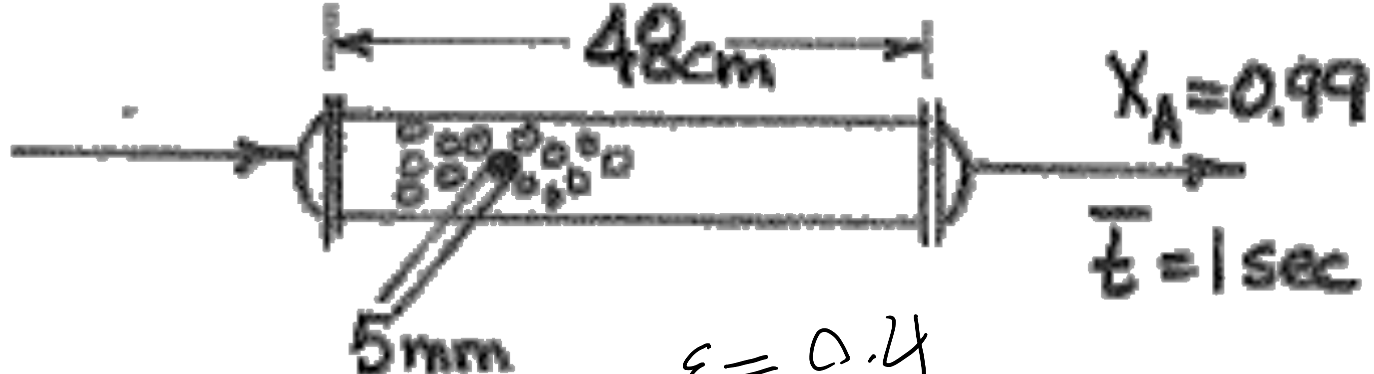
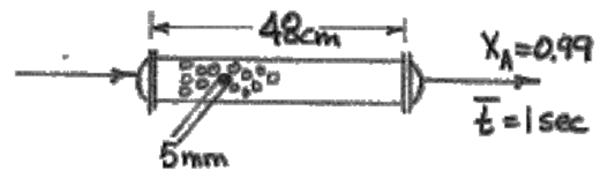


$\varepsilon = 0.4$

$Rep = 200$

13.11 First calculate k assuming plug flow, then account for dispersion

a). Find k assuming plug flow



For a 1st order reaction, $\varepsilon_A = 0$, we have

$$k\tau = \ln \frac{C_{A0}}{C_A} \quad \cdots \quad k_{plug} = \frac{1}{\tau} \ln \frac{C_{A0}}{C_A} = \frac{1}{(1\,sec)}(4.6) = 4.6\,sec^{-1} \qquad a)$$

13.11 b) Find k accounting for deviation from plug flow according to the
(continued) dispersion model

From the curve of Fig 17 we find for $Re_p = 200$ & $\epsilon = 0.4$

13.11 (continued) b) Find k accounting for deviation from plug flow according to the dispersion model

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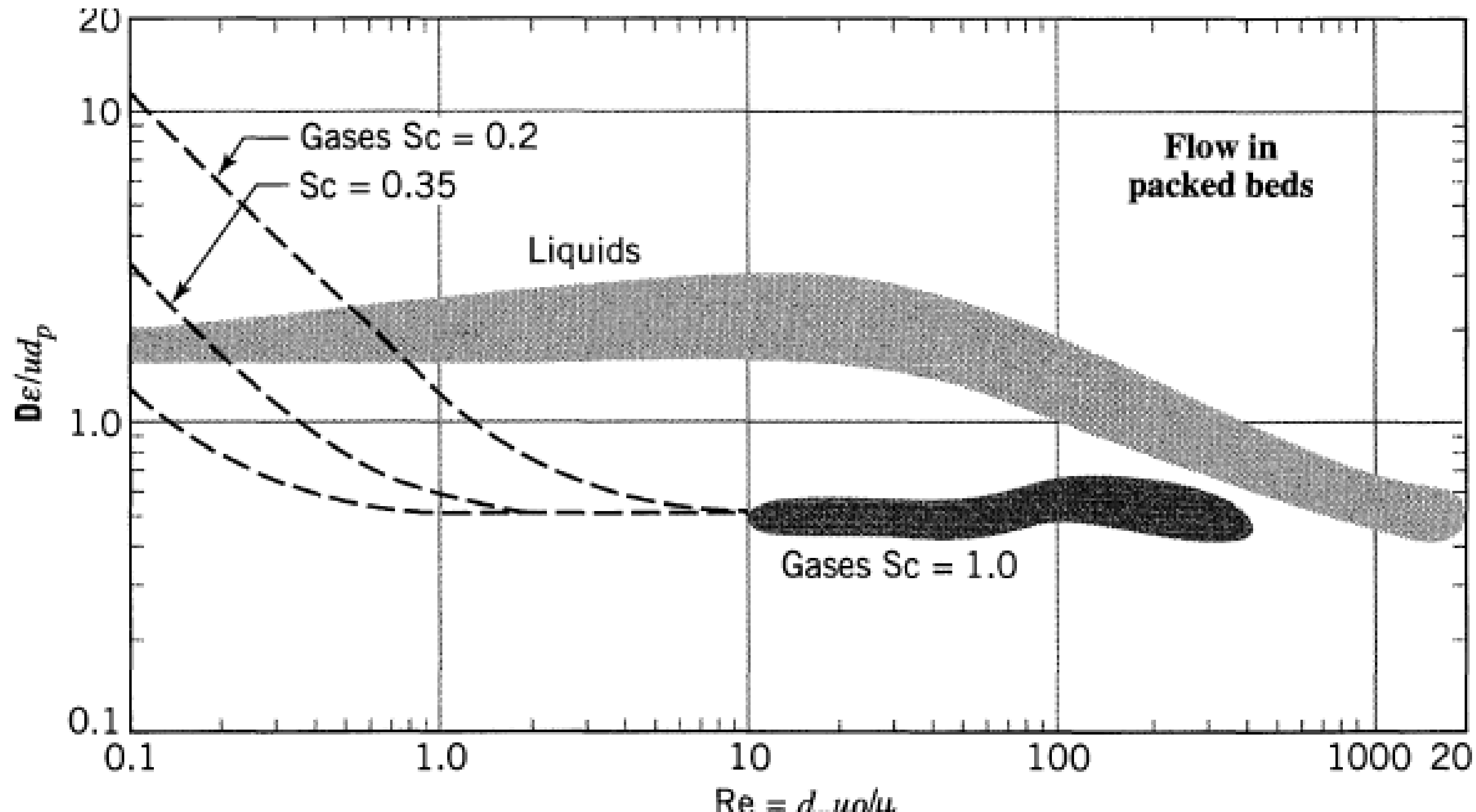

Figure 13.17 Experimental findings on dispersion of fluids flowing with mean ax velocity u in packed beds; prepared in part from Bischoff (1961).

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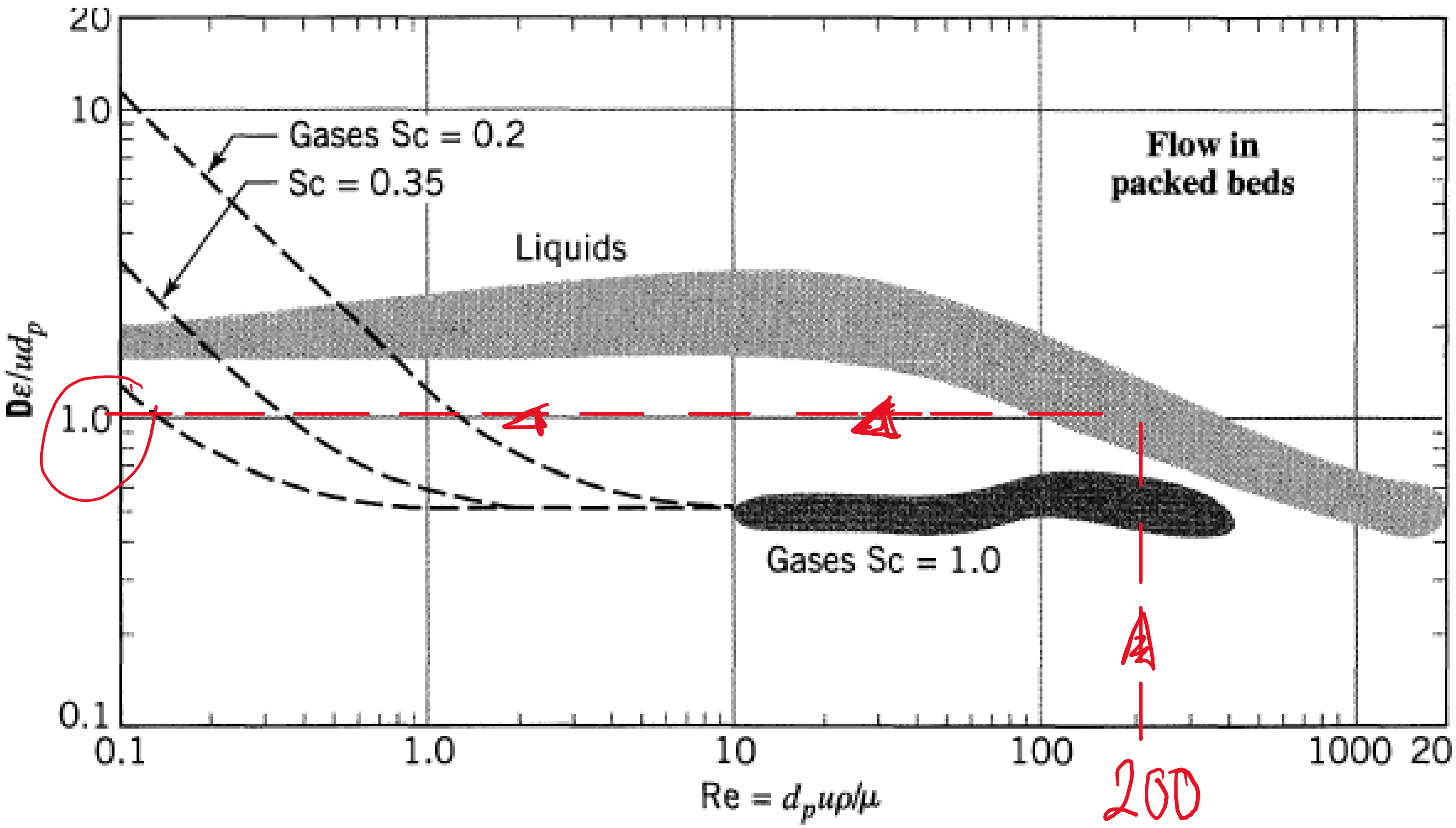
From the curve of Fig 17 we find for $Re_p = 200$ & $\epsilon = 0.4$



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From the curve of Fig 17 we find for $Re_p = 200$ & $\epsilon = 0.4$

$$\frac{D\epsilon}{ud_p} = 1 \qquad \text{... thus} \quad \frac{D}{ud_p} = 2.5$$

The size ratio of real to plug flow reactor, or what is equivalent, the ratio of the corresponding rate constants is found either from Fig 19 or from Eq 22 if the k ratio or the L ratio is close to unity. Use Eq. 22. Then

$$\frac{L}{L_p} = \frac{V}{V_p} = 1 + (k\tau)\frac{\mathbf{D}}{uL} \qquad \text{for same } C_{\text{A out}}$$

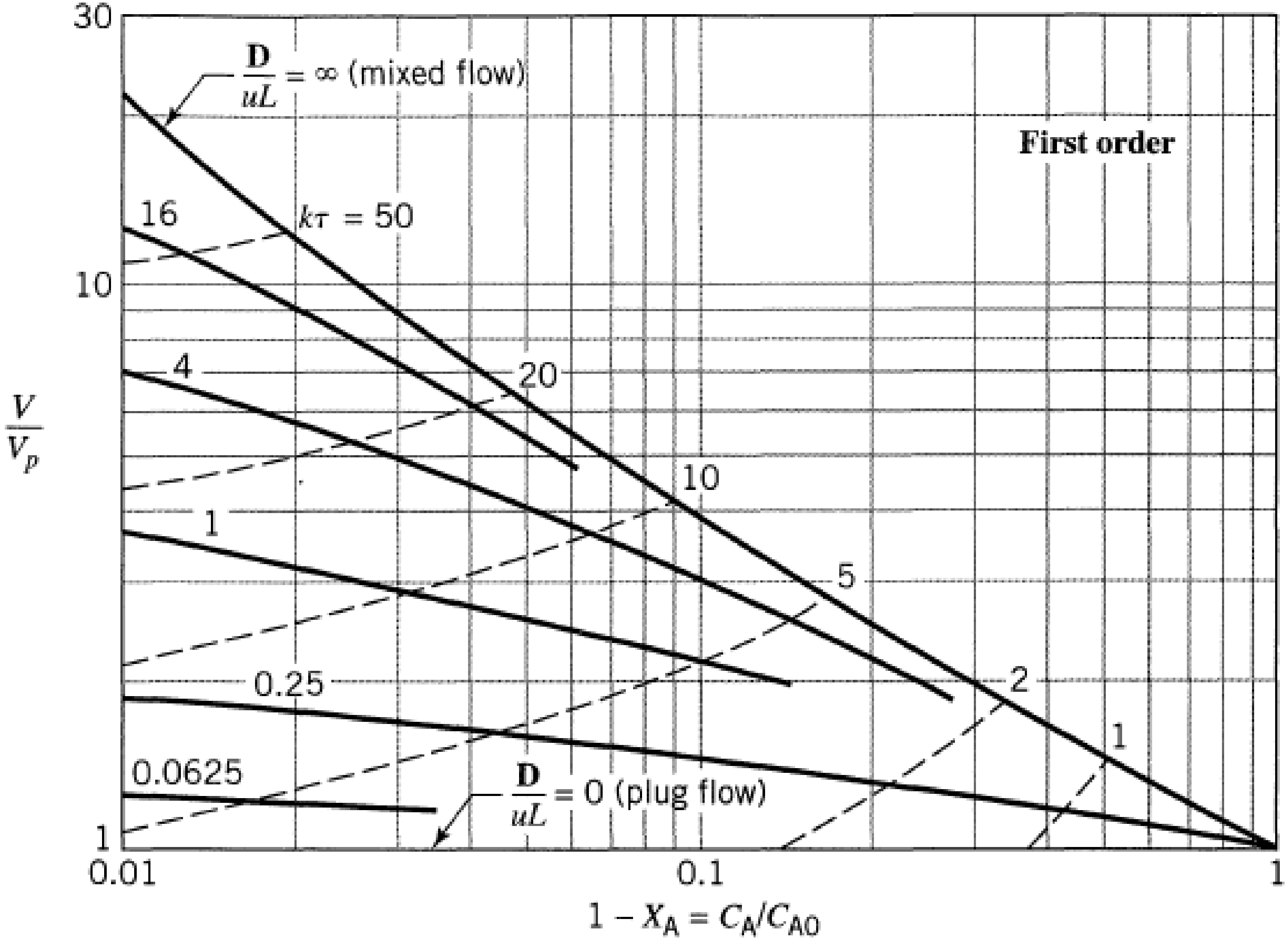

Figure 13.19 Comparison of real and plug flow reactors for the first-order A → products, assuming negligible expansion; from Levenspiel and Bischoff (1959, 1961).

Fig 19 or from Eq 22 if the $\bar{k}$ ratio or the L ratio is close to unity.
Use Eq. 22. Then

$$\frac{L}{L_p} = \frac{k_{true}}{k_{measured}} = 1 + k\tau\left(\frac{D}{uL}\right) = 1 + k\tau\left(\frac{D}{ud_p}\right)\left(\frac{d_p}{L}\right)$$

$$= 1 + 4.6(2.5)\left(\frac{5\,mm}{480\,mm}\right) = 1.1198$$

$$\therefore k_{true} = (4.6\,sec^{-1})(1.1198) = 5.15\,sec^{-1} \longleftarrow b)$$

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$$\therefore k_{true} = (4.6\,sec^{-1})(1.1198) = 5.15\,sec^{-1} \longleftarrow b)$$

Note: Direct use of Eq. 19 pg 314 (trial & error unfortunately) would give a more accurate answer.
Also, are you sure that $k_{true} > k_{plug}$? Convince yourself.

where

$$\boxed{\begin{aligned}\frac{C_A}{C_{A0}} = 1 - X_A &= \frac{4a \exp\left(\frac{1}{2}\frac{uL}{\mathbf{D}}\right)}{(1+a)^2 \exp\left(\frac{a}{2}\frac{uL}{\mathbf{D}}\right) - (1-a)^2 \exp\left(-\frac{a}{2}\frac{uL}{\mathbf{D}}\right)} \\ a &= \sqrt{1 + 4k\tau(\mathbf{D}/uL)}\end{aligned}} \qquad \textbf{(19)}$$

Figure 13.19 is a graphical representation of these results in useful form, prepared by combining Eq. 19 and Eq. 5.17, and allows comparison of reactor sizes for plug and dispersed plug flow.

Fig 19 or from Eq 22 if the $\bar{k}$ ratio or the L ratio is close to unity.
Use Eq. 22. Then

$$\frac{L}{L_p} = \frac{k_{true}}{k_{measured}} = 1 + k\tau\left(\frac{D}{uL}\right) = 1 + k\tau\left(\frac{D}{ud_p}\right)\left(\frac{d_p}{L}\right)$$

$$= 1 + 4.6(2.5)\left(\frac{5\,mm}{480\,mm}\right) = 1.1198$$

$$\therefore k_{true} = (4.6\,sec^{-1})(1.1198) = 5.15\,sec^{-1} \longleftarrow \text{b)}$$

Note: Direct use of Eq. 19 pg 314 (trial & error unfortunately) would give a more accurate answer.
Also, are you sure that $k_{true} > k_{plug}$? Convince yourself.

c) Error in assuming plug flow $= \left(\frac{5.15-4.60}{5.15}\right) \times 100 = 11\%$ underestimate $\longleftarrow$ c)