

DESIGN MANUAL

SPD0021ERN-USA

PROCEDURE DESIGN MANUAL

REACTOR

REV. : 2

DATE: 19 Jul 2011

HOLDS:

1. Section 3.4 on Nozzles – to be provided later



**SAMSUNG
ENGINEERING**

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1. SCOPE

This Basic Design Manual covers basic design considerations and calculation procedures for a fixed-bed reactor used in the petroleum industry for hydrotreating processes such as hydrodesulphurization and hydrocracking. For this design work, the following information must be available:

- 1) Heat and material balance around the reactor, including desired conversion, obtained by the reaction, and production capacity.
- 2) Catalyst volume required
- 3) Phases involved
- 4) Physical properties of the reactants and products.

2. FORMAT

In addition to the data sheets for pressure vessels, which indicate the general process requirements of the reactor, data sheets for internals must be provided. Data sheets for reactor internals should be prepared according to the instructions described in Section 3.5. Dimensions are to be shown on data sheet and should be in accordance with the descriptions in Section 3.3.

3. BASIC CONSIDERATIONS

This chapter describes the basic considerations concerning the bed dimensions and reactor internals for a fixed-bed reactor.

These considerations constitute the basis of the calculation procedures given in the next chapter.

3.1. Reactor Volume

The amount of catalyst required is specified by the process licensor or catalyst manufacturer. Sometimes, it is given in terms of space time(τ) or space velocity (S). These terms are defined as follows:

$$V = \tau \cdot F = \frac{F}{S} \quad (\text{Eq.3-1})$$

Where,

V: Reactor volume, m³ (ft³)

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F: Volumetric feed rate, m³/h (ft³/h)

τ : Residence time or space time, h

S: Space velocity, h⁻¹

The volumetric feed rate is usually at the condition of temperature and pressure, vapor or liquid, in the reactor. However, for gases, the space velocity is sometimes specified at standard temperature and pressure, in which case F must be calculated at 1 atmosphere and 0°C. When this is the case, since all gases occupy 22.4 liters/mol (~359 ft³/lb-mole) at these conditions, simply multiply the molar flow of reactants in gr-mole/h time 22.4 liters/mol to obtain F in liters/h.

Sometimes the space velocity is given as the liquid hourly space velocity. In this case the flow F is volumetric flow of liquid reactant, even when the reaction takes place in the vapor phase. This case F is obtained by dividing the reactant mass flow by its liquid density.

The relation between the space-velocity and space-time for actual entering conditions under standard conditions is given by:

$$t' = \frac{1}{S'} = t \frac{C'A_0}{CA_0} = \frac{1}{S} \frac{C'A_0}{CA_0} \quad (\text{Eq.3-2})$$

Where,

t: Space time under standard condition

t': Space time at actual inlet condition

S: Space velocity under standard condition

S': Space velocity at actual inlet condition

CA₀: Molar concentration of component A in the reactant at standard conditions

C'A₀: Molar concentration of component A in the reactant at actual inlet conditions.

In addition to the volume occupied by the catalyst, space must be provided for reactor internals, inert balls above and below the catalyst bed.

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3.2. Bed Dimensions

When the catalyst volume has been set, bed diameter and height may be determined. The bed geometry is often indicated as bed L/D ratio, but the bed dimension is not determined solely by bed L/D. It is also based on good contact of reactants with catalyst and economic considerations.

Mass velocities through the bed must be high enough to assure good contact (minimizing interface gradients) and good distribution.

Increases in pressure drop, however, may affect pumping or compressing costs. These costs should not exceed the saving realized through improved reactor performance at high mass velocities. Thus, the bed L/D ratio should be based on economics and reactor performance.

In many fixed-bed systems, the most economical pressure drop is in the range of 3% to 15% of the operating pressure. Typical examples drawn from recent designs are listed in Table 3-1 for guidance.

Criteria for determination of the bed dimension are summarized as follows:

- To minimize interface gradients the bed Reynold's number should be greater than 100. The bed Reynold's number is defined by equation 4-5 in Section 4.2.
- To obtain good distribution of liquid in a trickle-bed reactor, in which liquid and gas flow counter-currently, the liquid mass velocity should exceed 20 m/hr. To eliminate wall effect in a trickle-bed reactor, the bed L/D ratio per bed should be lower than 3.
- To prevent catalyst attrition, the design load imposed on the catalyst particles at the bottom of the bed should not exceed the crush strength of the catalyst.
- To avoid bed lifting during up-flow operation, the bed pressure drop should be kept below the bed lifting pressure drop, preferably less than 50%. In any case it should not be greater than 75% of bed lifting pressure drop for initial design.

The procedure for calculation of the bed pressure drop is described in Section 4.3.

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3.3. Internals

Poor distribution of reactants in a fixed bed reactor will decrease the yield and conversion of the reaction. To prevent maldistribution, reactors should be equipped with certain kinds of internals such as inlet distributors, liquid distributor trays and collector rings. Typical arrangements of these internals are shown in Figs. 3-1 and 3-2.

The internals indicated in Fig. 3-1 must be used for a vapor phase reactor, and those in Fig. 3-2, for a mixed phase reactor.

3.3.1. Inlet Distributor

The main reason for installing an inlet distributor is to prevent a cavity on the surface of the catalyst bed through the jet action of the feed gas stream.

An inert ball layer above the catalyst bed can smooth out the feed gas flow better than an inlet distributor does, and a properly designed liquid distributor tray can distribute the liquid flow more uniformly over the entire cross section of the bed than an inlet distributor.

Two types of inlet distributors are available and the selection is made according to the phase of the fluid handled.

A standard design slotted distributor for an all-vapor phase, is shown in Fig. 3-3.

A standard design impingement baffle for a mixed phase is shown in Fig. 3-4.

3.3.2. Liquid Distributor Tray

A good liquid distribution over the catalyst bed is essential for obtaining the desired conversion and for reducing side reactions such as coking. For this purpose, a liquid distributor tray must be installed over the catalyst bed. The distributor consists of a flat plate and a number of chimneys. The chimney is located on a triangular pitch with 150 mm spacing, as shown in Fig. 3-5 as a standard design. Liquid is distributed uniformly across the bed through holes drilled on the sidewall of the chimney.

The hole diameter should be selected to hold the liquid layer over the side holes specified as 'h' in Table 3-2, even if it is operated at minimum liquid load.

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The height of the chimney above the holes, c , depends on the range of turndown required. If the load of the reactor varies between 50% and full design load, the height of the chimney above the holes, c , will be as show in Table 3-2.

3.3.3. Baskets

The baskets should be placed on the top of the catalyst bed to catch any scale carried with the feed, and to maintain the surface area of the catalyst bed by preventing it from plugging with deposits.

The baskets are 150 mm in diameter and 500 mm long, as shown in Fig. 3-6 and located on a 250 mm concentric circular pitch. The number of baskets depends on the inside diameter of the reactor, as shown in Fig. 3-6.

3.3.4. Inert Balls

Inert balls must be provided above and below the catalyst bed as shown in Figs. 3-1 and 3-2. The pressure drop through the inert balls can be calculated by the equation described in Section 4.1. However, this pressure drop is usually so small, compared to the pressure drop through the catalyst bed, that it is generally ignored.

A 150 mm layer of 15 mm inert balls is placed on top of the catalyst bed. The bottom of the reactor must be filled with three layers of inert balls, to prevent the catalyst from penetrating and escaping the bed. The height and size of inert balls of each layer are shown in Figs. 3-1 and 3-2. The depth of each layer depends on the geometry of the reactor head but should be larger than 75 mm, preferably 150 mm.

The top layer, 75 mm layer of 6 mm inert balls is placed directly under the catalyst bed, to prevent catalyst penetration. The size of 6 mm is selected on the basis of preventing 1.5 mm catalyst penetration. Typical characteristics of inert balls are summarized in Table 3-3.

3.3.5. Collector Ring

At the bottom of the reactor, a collector ring must be installed upstream of the reactor outlet to prevent the catalyst from escaping with products which will cause mal-distribution in the catalyst bed. The standard design for the collector ring is shown in Fig. 3-7.

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The diameter of the collector ring should be roughly one-third the reactor diameter ($1/3 D$). The top dished head should be perforated over its entire surface with 20mm holes on a 50mm triangular pitch.

The length of slots must be roughly 6.5% of reactor diameter, but not less than 80 mm.

3.4. Nozzles

3.4.1. Inlet Nozzle

3.4.2. Outlet Nozzle

3.4.3. Thermowell Nozzle

3.4.4. Catalyst Unloading Nozzle

This section will be provided later on.

3.5. Design Conditions

For mechanical design, the following design data should be provided on the data sheet.

3.5.1. Catalyst Bed

- Design pressure drop of the catalyst bed should be double the calculated bed pressure drop obtained by Eq. 4-4 in Section 4.1, to allow for differences in catalyst loading and for the possible buildup of coke.
- This will usually cover the significant bed pressure drop due to the high velocity of the process stream during depressuring of the reactor.
- The design pressure drop should be the sum of the design pressure drop of the catalyst bed and the catalyst support device.
- The liquid density should be that of the heaviest liquid introduced to the reactor in any operation.
- The packed and true density of a catalyst should be weighted by multiplying with the correction factor shown in Table 3-4.
- The true density of a catalyst or inert ball is provided to predict liquid holdup. A void fraction of the bed can be used for this purpose.

3.5.2. Liquid Distributor Tray

- The liquid density should be that of the heaviest oil. Sludge density should be indicated, accompanied with the liquid density in parenthesis. The sludge density is usually around 3,000 kg/m³.

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- The design pressure drop across the liquid distributor tray should be double the calculated pressure drop.
- Maximum allowable tilt: See Table 3-5
- Maximum allowable deflection shall not be more than the value given in Table 3-6.

3.5.3. Inlet Distributor

Design pressure drop of the inlet distributor should be double the calculated pressure drops.

4. CALCULATION PROCEDURE

4.1. Bed Pressure Drop

An outline for the calculation of bed pressure drop is given below.

- 1) Calculate the single-phase friction losses, for the gas and the liquid respectively, from the Ergun equation.

In its simplest form, the Ergun equation represents a bed friction factor as a function of its Reynolds number.

$$f_p = \frac{150}{Re_p} + 1.75$$

Where

$$f_p = \frac{\Delta P D_p}{L \rho U_m^2} \frac{\epsilon^3}{1-\epsilon}$$

and

$$Re_p = \frac{D_p U_m \rho}{(1-\epsilon)\mu}$$

Substituting, this equation becomes:

$$\Delta P g_c = \frac{150 (1-\epsilon)^2 \mu U_m}{\epsilon^3 D_p^2} + \frac{1.75 (1-\epsilon) G U_m}{\epsilon^3 D_p}$$

Reference: *Chemical Engineering Progress*, Feb 1952, Vol. 48, No.2
Fluid Flow through Packed Columns, Sabri Ergun

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Where,

ΔP : Pressure drop per unit length

g_c : Gravitational constant

ϵ : Fractional voids in packing, void volume/unit volume of column
(This value is usually in the range of 0.38 to 0.40 for freshly-packed beds, and 0.36 to 0.38 for used beds. For other particles, Fig. 4-1 is provided as a guide)

μ : Viscosity

U_m : Superficial velocity (based on empty column cross section)
measured at average pressure which must be assumed and then confirmed by calculation, for the gas

D_p : Effective particle diameter. Diameter of a sphere with a surface area equivalent to one catalyst particle

$$D_p = \frac{6(1-\epsilon)}{S}$$

(Eq. 4-1b)

S : Surface area of packing per unit volume

G : Mass velocity based on unpacked flow area $= \rho U_m$

ρ : density

2) Calculate the value of X from Eq. 4-2

$$X = \frac{\Delta P_l}{\Delta P_g}$$

(Eq. 4-2)

3) Calculate the two-phase pressure drop (ΔP_{lg}) from Eqs. 4-3 and 4-4.

$$\log Y = \frac{0.416}{(\log x)^2 + 0.666}$$

(Eq. 4-3)

$$\Delta P_{lg} = Y(\Delta P_l + \Delta P_g) \quad (\text{Eq. 4-4})$$

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The two-phase pressure drop can then be calculated with the aid of the single-phase losses. The value of Y usually varies from 4.2 to 1.5.

4.2. Reynolds Number and Friction Factor

The Reynolds number Re is defined as

$$R = \frac{D_p G}{\mu}$$

(Eq. 4-5)

Where,

G: Fluid superficial mass velocity (kg/sec.m²)

μ: Fluid viscosity (kg/m.sec)

D_p: Particle diameter (m)

For non-spherical particles,

$$D_p = \frac{6 \cdot (1 - E)}{\phi \cdot S}$$

(Eq. 4-6)

Where,

S: Specific surface, or area of particle surface per unit volume of bed = S_o (1-E), (m²/m³)

S_o: Area of particle surface per unit volume of solids (m²/m³)

φ: Shape factor, are tabulated in Table 4-1

E: Actual voidage of the bed, is shown in Fig.4-1 (as a guide only)

The friction factor and the flow pattern, functions of the Reynolds number, are given in Fig. 4-2.

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4.3. Bed Lifting

For upflow operation, when the pressure drop just equals the weight of the catalyst, the particles will just begin to be in motion. This will be the onset of bed lifting. So, the pressure drop must be kept below the bed lifting pressure drop; preferably less than 50% of this value.

$$\Delta P_E = \rho_c - (1 - E)\rho_l$$

(Eq. 4-7)

Where,

ΔP_B : Bed lifting pressure drop per unit length (kg/m².m)

ρ_c : Bulk density of catalyst (kg/m³)

E: Void fraction of bed (dimensionless)

ρ_l : Liquid density at conditions (kg/m³)

4.4. Inlet Distributor

Pressure drop across the inlet distributor is the sum of ΔP_1 , ΔP_2 and ΔP_3 , as calculated by Eqs.4-8 to 4-10.

$$\Delta P_1 = \frac{\rho_g \cdot (U_l - U_i)^2}{10^4 \times 2g_c} \quad (\text{Eq. 4-8})$$

$$\Delta P_2 = \frac{1.3 \cdot \rho_g \cdot U_i^2}{10^4 \times 2g_c} \quad (\text{Eq. 4-9})$$

$$\Delta P_3 = \frac{2.8 \cdot \rho_g \cdot U_s^2}{10^4 \times 2g_c} \quad (\text{Eq. 4-10})$$

Where,

ΔP_1 : Pressure drop due to expansion (kg/cm²)

ΔP_2 : Pressure drop due to impingement of gas on bottom plate of inlet distributor (kg/cm²)

ΔP_3 : Pressure drop through the slots, (kg/cm²)

ρ_g : Gas density at conditions, (kg/m³)

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- U_i: Gas velocity in external piping (m/sec)
- U_i: Gas velocity in inlet distributor. (m/sec)
- U_S: Gas velocity in slots (m/sec)
- g_C: Gravitational constant, 9.81 m/sec² (32.17 ft/sec²)

4.5. Liquid Distributor Tray

The number of chimneys located on a 150 mm triangular pitch can be estimated by the following Equation,

$$N = \frac{(\text{Cross Sectional Area of Tower}) \times 0.8}{0.866} \quad (\text{Eq. 4-11})$$

The hole diameter is calculated from the following Equation,

$$h^4 = \frac{16.76 \times 10^6}{h_m} \left(\frac{L_m}{N} \right)^2 \quad (\text{Eq. 4-12})$$

Where,

- L_m : Minimum liquid load (m³/hr)
- N : Number of holes (dimensionless)
- h_m : Minimum liquid height over the holes (mm) shown in Table 3-2.
- h : Hole diameter (mm)

For the pressure drop across the chimney, the following equation is set up,

$$\Delta P = 7.66 \times 10^{-6} \rho_g U^2 \quad (\text{Eq. 4-13})$$

Where,

- ΔP: Pressure drop (kg-force/cm²)
- U : Gas velocity in chimney (m/sec)

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ρ_g : Vapor density at conditions (kg/m³)

4.6. Collector Ring

The pressure drop through the collector ring and outlet nozzle is the sum of ΔP_C and ΔP_n , as calculated from Eqs. 4-14 and 4-15.

$$\Delta P_c = \frac{2.8 \cdot \rho_g \cdot U_s^2}{10^4 \times 2 \cdot g_c}$$

(Eq. 4-14)

Where,

ΔP_C : Pressure drop through holes and slots of collector (kg/cm²)
($\geq P$ through the screen is ignored),

C : Vapor density at conditions (kg/m³)

U_s : Vapor velocity through holes and slots (m/sec)

g_c : Gravitational constant, 9.81 m/sec² (32.17 ft/sec²)

$$\Delta P_n = \frac{1}{2} \frac{\rho_g \cdot U_l^2}{10 \times 2 \cdot g_c}$$

(Eq. 4-15)

Where,

ΔP_n : Pressure drop due to sudden contraction at entrance to the outlet nozzle (kg/cm²)

U_l : Vapor velocity in outlet nozzle (m/sec)

ρ_g : Vapor density at conditions (kg/m³)

g_c : Gravitational constant, 9.81 m/sec² (32.17 ft/sec²)

4.7. Liquid Saturation

Liquid saturation (R_l) is defined as the value which is obtained by dividing the liquid holdup by the total void volume in the packed bed.

$$V_{\text{liquid}} = (V_{\text{bed}}) (E) (R_l)$$

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(Eq. 4-16)

The liquid saturation (R_ℓ) can be obtained from Eq. 4-17.

$$\text{Log} (R_\ell) = -0.774 + 0.525(\log X) - 0.109(\log X)^2 \quad (\text{Eq. 4-17})$$

Where,

R_ℓ : Liquid saturation (dimensionless)

X : Variable defined by Eq. 4-2

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5. REVISION HISTORY

SEA issue of SPD0021ERN-USA

REV NO.	DATE	SCOPE	DESCRIPTION
2	21 Dec 2010	Revised for new Doc. Numbering System	-Team: Process Engineering -Prepared by: K. Elliott -Checked by: S. Wenger -Approved by: M. Czaikowski

Previously issued of SEM-9417E-USA

REV NO.	DATE	SCOPE	DESCRIPTION
1	01 Dec 2010	All Pages	First Edition—SEA -Team: Process Engineering -Prepared by: C. Vasquez / P.N. Reddy -Checked by: M. Czaikowski / H.M. Feintuch -Approved by: J. Tait

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6. APPENDIX A. Tables and Figures

Table 3-1: Examples of Reactor Design

Mixed-phase service	A	B (ARD)	C (KD)
Bed L/D Ratio (-)	3.5	3.93	2.15
Pressure drop through reactor (kg/cm ²)	0.63 (cal)	2.53 (Design)	3.0 (Design)
Mass liquid velocity (m/hr)	20	18	38

Table 3-2: Height of Liquid Layer and Chimney

(See Fig. 3-5 also)

Tower Dia. (mm)	~ 2500	2500 - 4000	4000 - 6000
h: hole dia. (mm)	25	35	45
c: chimney height (mm)	120	170	220

Table 3-3: Typical Characteristics of Inert Balls

	Type	Ceramics	Mullite	Alumina
		60% SiO ₂	20% SiO ₂	0.05% SiO ₂
		27% Al ₂ O ₃	67% Al ₂ O ₃	99% Al ₂ O ₃
Bulk Density	(kg/m ³)	1400	1700	2000
True Density	(kg/m ³)	2250	3000	3400
Crush strength	(kg/cm ²)	2000	4500	2500

Table 3-4: Correction Factor for Packed and True Density

Feed Fluid	Spherical	Extrudate
Gas oil or VOG	1.2	1.5
Reduced Crude	1.5	2.0

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Table 3-5: Maximum Allowable Tilt

Tower Dia.	Tolerance
Not more than 1500 mm	3 mm
1,501 mm - 2,500 mm	4.5 mm
2,501 mm - 4,000 mm	6 mm
4,001 mm - 6,000 mm	9 mm
6,001 mm - 8,000 mm	12 mm
8,001 mm - 10,000 mm	15 mm

Table 3-6: Maximum Allowable Deflection

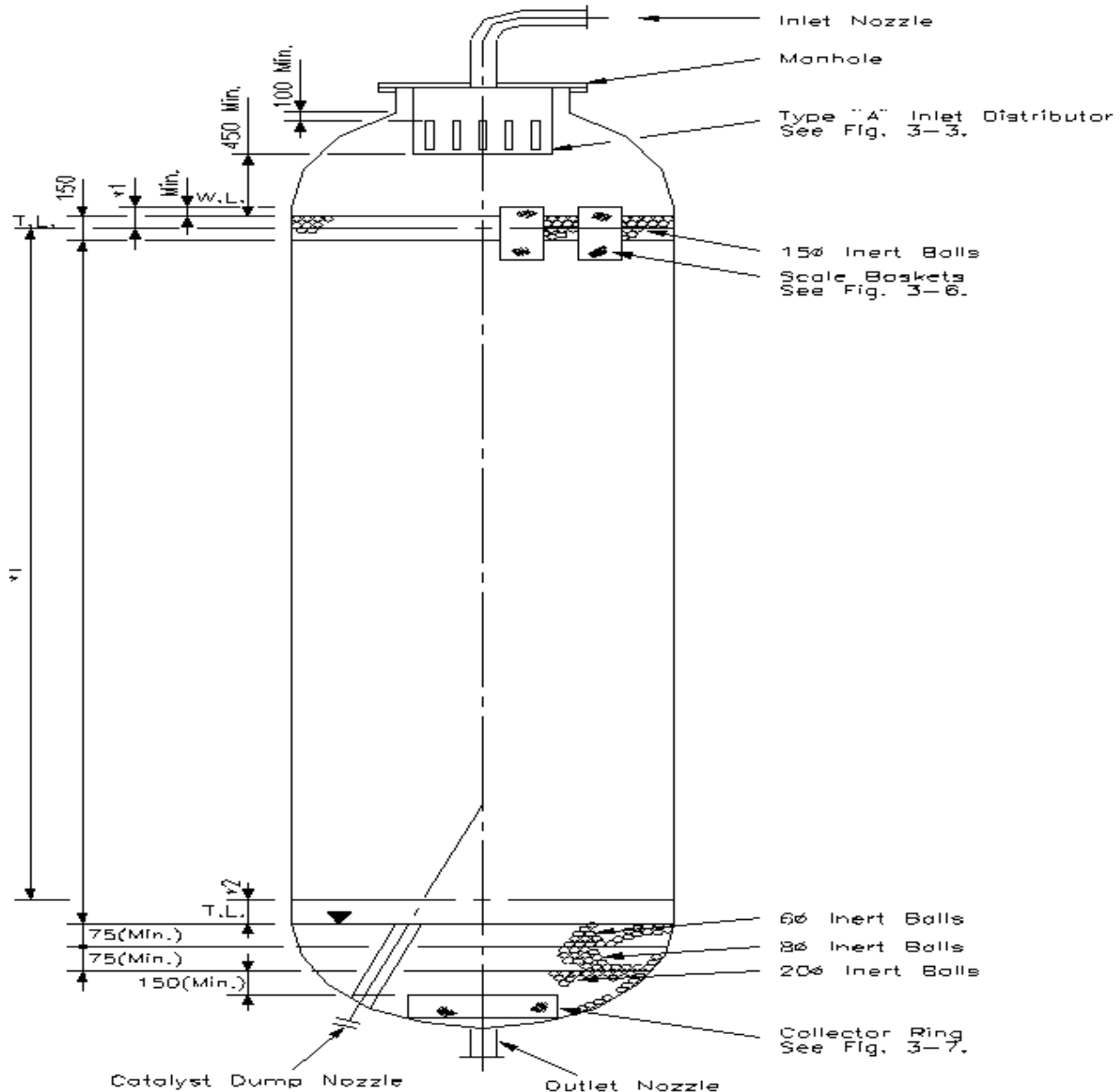
Tower Dia.	Max. Allowable Deflection
Less than 2,200 mm	3 mm
More than 2,201 mm	1/720 of the tower dia.

Table 4-1: Shape Factor (Shape Factors for Non-Spherical Particles)

Material	Nature of Grain	Φ
Arnold's wire spirals		0.2
Berl saddles		0.3
Raschig rings		0.3
Cool dust, natural		0.65
Cork 0.69		0.69
Flue dust	Fused, spherical	0.89
Flue dust	Fused, aggregated	0.55
Glares, crushed	Jagged	0.65
Sand	Rounded	0.83
Sand	Angular	0.73

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Fig. 3-1: Typical Fixed-Bed Reactor for Vapor-Phase Service

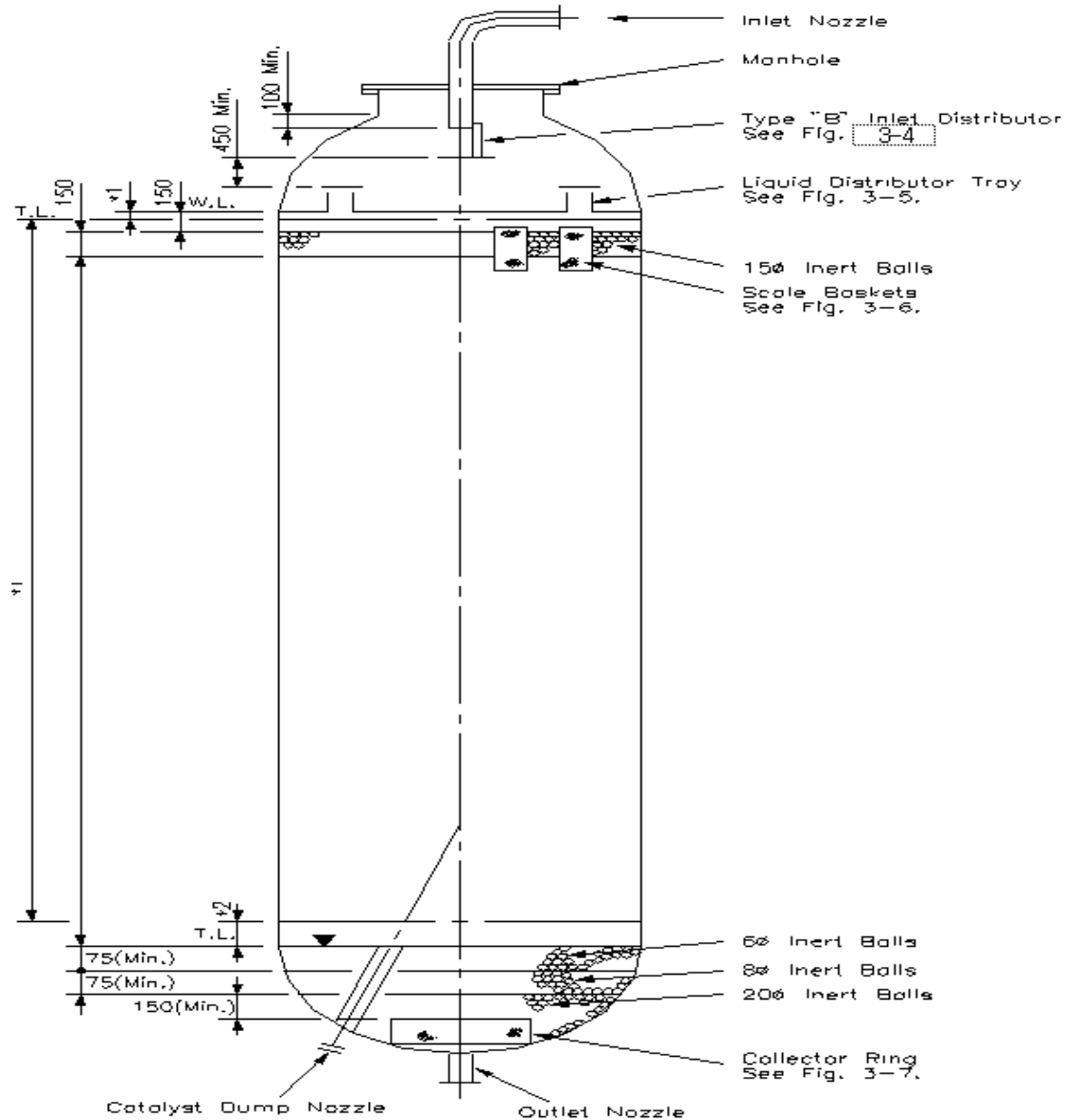


Notes:

- *1. To be decided by mechanical designer
- *2. To be decided by mechanical designer taking into account inside diameter shown by mark(▼), which is not less than 95% of nominal shell inside diameter.
- 3. Elevation above grade to be determined by details of catalyst dumping procedure.

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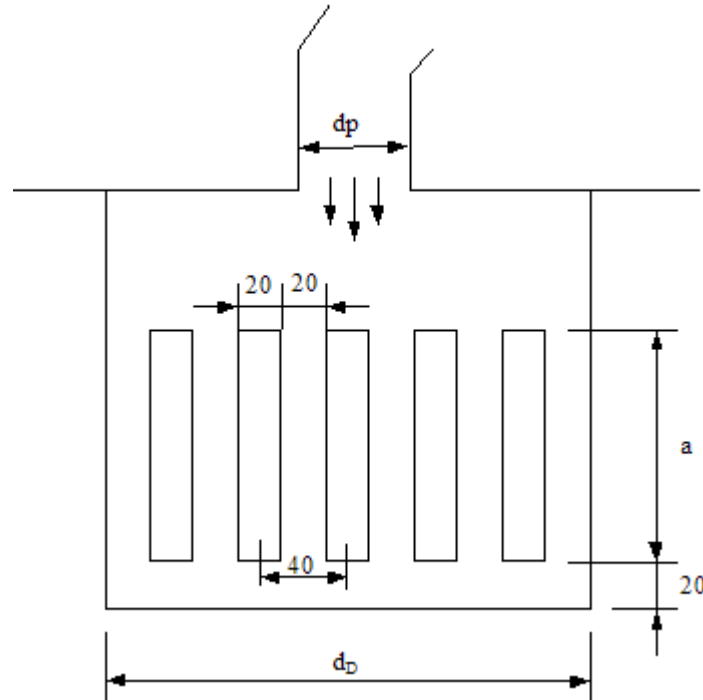
Fig. 3-2: Typical Fixed-Bed Reactor for Mixed-Phase Service



Notes:

- *1. To be decided by mechanical designer
- *2. To be decided by mechanical designer taking into account inside diameter shown by mark(▼), which is not less than 95% of nominal shell inside diameter.
- 3. Elevation above grade to be determined by details of catalyst dumping procedure.

Fig. 3-3: Distributor Type A : Slotted Distributor



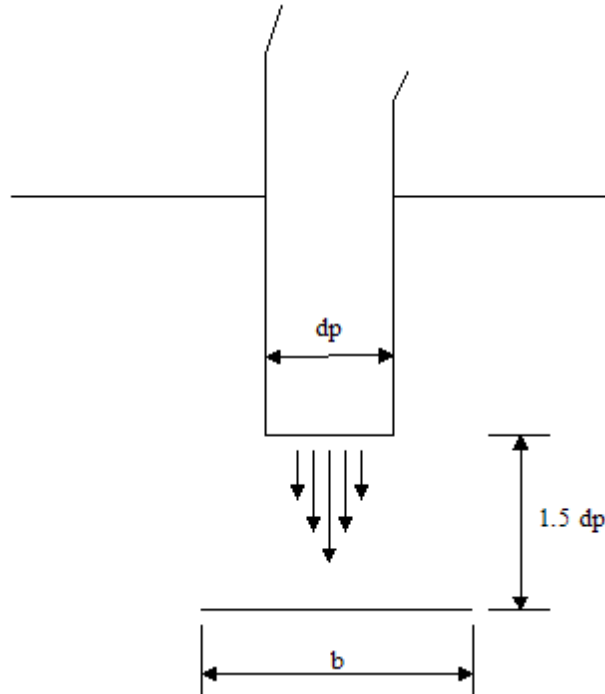
Distributor Diameter , d_D

Minimum Diameter = $1.2 d_p$

Maximum Diameter = Manhole diameter less 100 mm clearance

Set length "a" such that the slot area is three times the cross-sectional area of the inlet distributor. However, "a" must be equal to or larger than 100 mm.

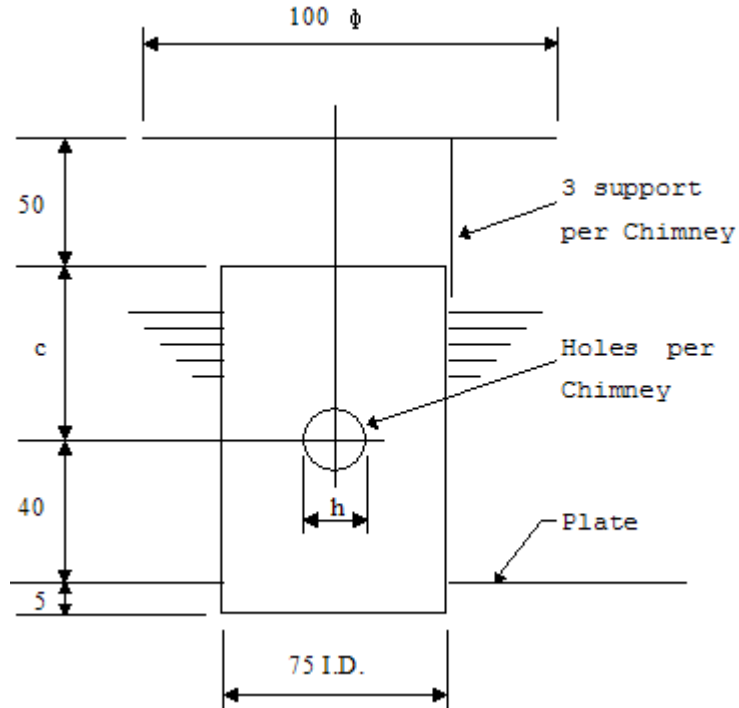
Fig. 3-4: Distributor Typer B : Impingement Baffle



Diameter Impingement baffle " b " should be larger than 1.2 times the inlet pipe diameter " dp " but should not exceed 2 times the dp or inside diameter of manhole.

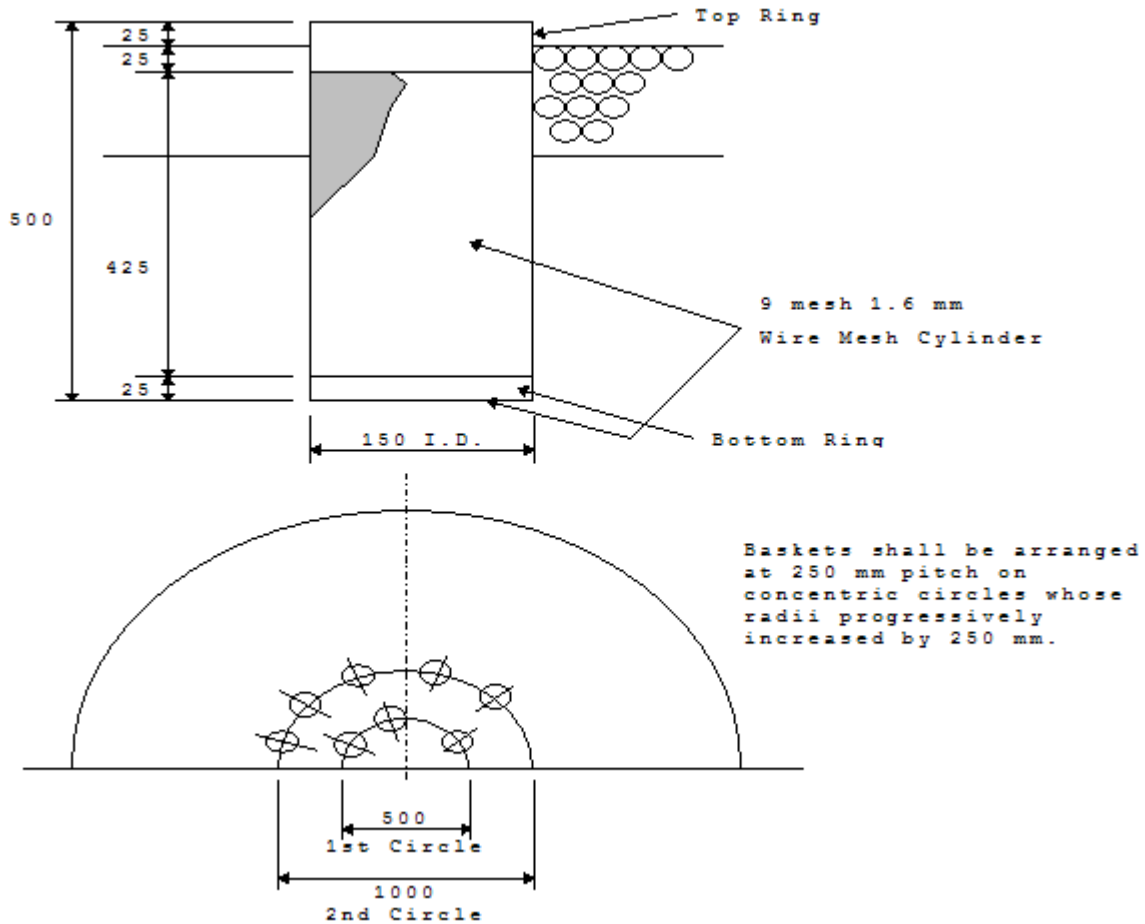
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Fig. 3-5: Chimney of Liquid Distributor Tray



Chimney shall be arranged on 150 mm triangular pitch for each tray plate

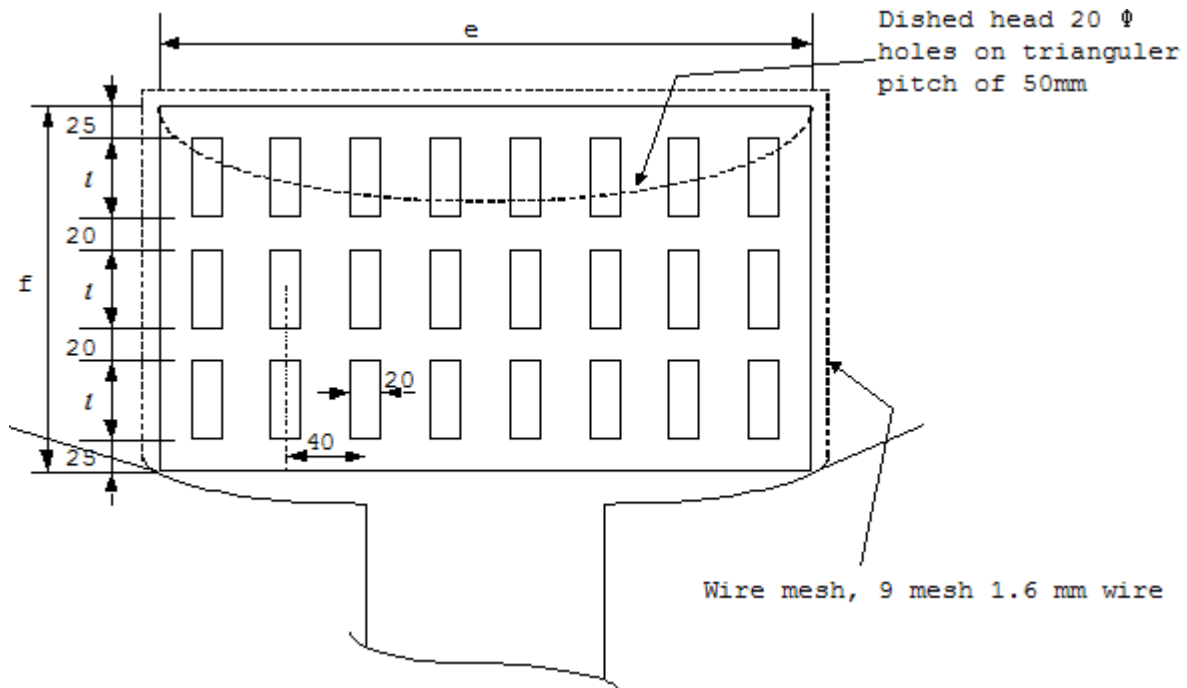
Fig. 3-6: Baskets



No. of Baskets

Circle	Circle Dia.	No. of Baskets
1st	500	6
2nd	1,000	12
3rd	1,500	18
4th	2,000	25
5th	2,500	31
6th	3,000	37
7th	3,500	44
8th	4,000	50
9th	4,500	56
10th	5,000	62

Fig. 3-7: Collector Ring

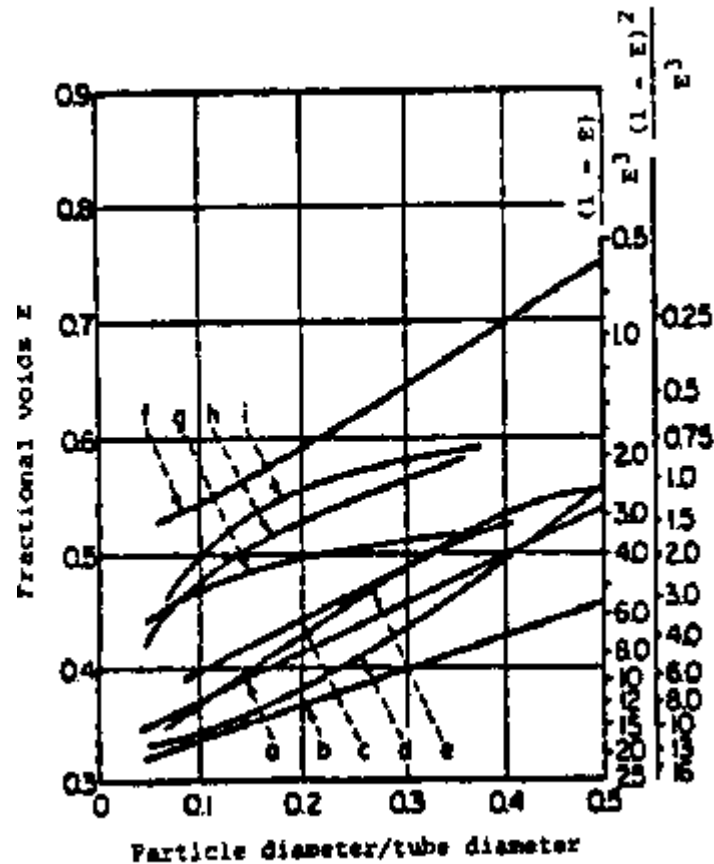


$$e = \frac{1}{3} D \text{ (Not less than 80 mm)}$$

$$l = 0.065 D$$

$$f = 3l + 90$$

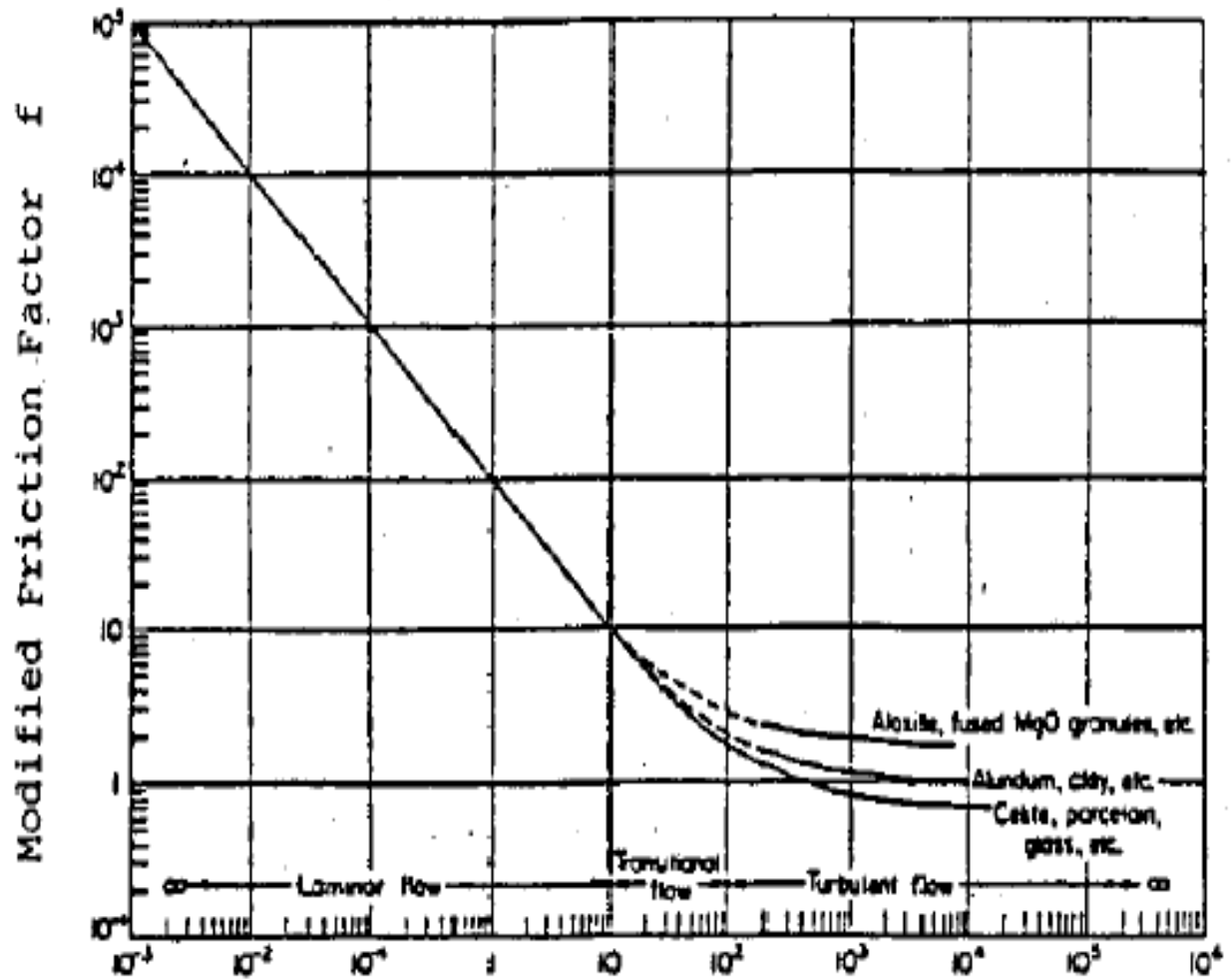
Fig. 4-1: Voidage in Packed Beds



- Spherical: a, smooth, uniform
 b, smooth,
 c, clay
- Cylindrical : d, smooth, uniform
 e, alundum, uniform
 f, clay Raschig rings
- Granules: g, fused magnetite (synthetic ammonia catalyst)
 h, fused alundum
 i, Aloxite

DESIGN MANUAL
REACTOR

Fig. 4-2: Friction Factor and Flow Pattern for Packed Beds of Solid



$$\text{Modified Reynolds } Re = \frac{D_p G}{\mu}$$