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Revision Memo

- 12/01** This is a minor revision. It contains some additional references and clarifications for a number of design considerations. The Parallel Flow method for more accurate pressure drop prediction is also introduced.



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SCOPE

This section gives recommended process and mechanical design procedures for single-phase (all-vapor or all-liquid) axial fixed bed reactors and for single-phase vapor radial flow reactors and tubular reactors. The direction of flow in an axial bed can be either upward or downward. In a radial flow reactor, flow is either directed radially inward or outward. Detailed mechanical considerations, shrouds, linings, and corrosion problems are covered in [Section VII-D](#), and special requirements of heavy-wall vessels are covered in [Section VII-E](#).

Operating conditions, the amount and type of catalyst required, and other process design information are contained in the Design Basis Specification (DBS). For detailed information on mechanical requirements, see Global Practices **GP 5-1-1**, **GP 5-1-2**, and **GP 5-2-2**.

REFERENCES

GLOBAL PRACTICES

- GP 5-1-1** Pressure Vessels
GP 5-1-2 Additional Requirements for Heavy Wall Pressure Vessels, Thickness Over 2 In.
GP 5-2-2 Internals for Fixed Bed Reactors

OTHER LITERATURE

1. *Design Guidelines for Radial Flow Fixed Bed Reactors*, ER&E Report No. EE.20E.76, November, 1976.
2. Ergun, *Fluid Flow through Packed Columns*, Chem. Eng. Prog. 48: 89-94 (Feb., 1952).
3. *Design Guidelines for Fixed Bed Vapor Phase Inlet Distributors*, ER&E Report No. EE.3E.74, January, 1974.
4. *Design Guidelines for Maximum Allowable Pressure Drop in Fixed Beds*, ER&E Report No. EE.111E.74, December, 1974.
5. *Radial Reactor Design*, Deck No. 3695, ER&E Report No. EE.41E.77, March 1977.
6. *Computer Model for Tubular Fixed Bed Reactors with Arbitrary Kinetics* - Deck No. D9317, ER&E Report No. EE.100E.82, December, 1982.
7. *Experimental Evaluation of Fixed Bed Reactor Inlet Nozzle Distributors*, DAN 98-M-0358, April, 1998.
8. *An Improved Method for Predicting Two-Phase Pressure Drop in Packed Beds*, 96M-0654, December, 1996
9. *Implementation of the Parallel Flow Method for Prediction of Two-Phase Pressure Drop in Fixed Beds*, 97M-1152, December 1997.

DEFINITIONS

See [Section VII-A](#).

APPLICATIONS

The following processes or services are examples of applications in which single-phase fixed bed, radial, and tubular reactors are used:

- Drying
- Hydrofining
- Methanation (hydrogen plants)
- Phthalic anhydride
- POWERFORMING
- Shift conversion (hydrogen or synthesis gas plants)
- Vapor phase hydrogenation (e.g., acetylene converters)
- Zinc oxide reactors (sulfur removal)



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BASIC DESIGN CONSIDERATIONS

FIXED BEDS

Fluid Flow and Contacting

Distribution of the process fluid and its contacting with the catalyst are affected by inlet flow conditions, design of the inlet distributor, and the mass velocity (pressure drop) in the reactor. If uniform flow is to be obtained in the bed, the dynamic head $\rho u^2/2$ (where ρ and u are the fluid density and velocity) of the incoming stream must be small compared to the resistance through the bed. Therefore, the ratio of dynamic head to total bed pressure drop (expressed in consistent units) is a key parameter for determining flow uniformity in the bed. The allowable ratio of dynamic head of the inlet stream to the bed pressure drop depends on the distributor design. This is discussed below for various inlet distributor designs.

To ensure uniform contacting in the catalyst bed, the flow in the bed should be in the "turbulent" regime. In situations where this may not be feasible, as in the regeneration cycles of POWERFORMING units and HYDROFINING units it is desirable at least to keep the flow out of the laminar regime. The Reynolds number, which is a measure of the flow regime in the bed, is the major criterion for good catalyst contacting. Required Reynolds numbers for various operations are given below.

Minimum Bed Reynolds Number for Uniform Contacting - For all designs, the bed Reynolds number should be maintained greater than 100 for reaction. For regeneration, where this may not be feasible, a lower limit of 20 for the Reynolds number is recommended:

S For reaction:
$$Re = \frac{W D_p}{\mu (1 - \epsilon)} > 100 \quad \text{Eq. (1)}$$

S or regeneration:
$$Re = \frac{W D_p}{\mu (1 - \epsilon)} > 20 \quad \text{Eq. (2)}$$

where: Re = Reynolds number, dimensionless
 W = Mass velocity of vapor or liquid, lb/hr ft² (kg/sm²) based on reactor cross sectional area
 D_p = Equivalent particle diameter, ft (m) (see [Section VII-A](#))
 μ = Viscosity of vapor or liquid at conditions, lb/ft hr (kg/ms)
 ϵ = Bed void fraction, dimensionless. See [Table 1](#) of [Section VII-A](#) for typical values.

Minimum Bed Pressure Drop - In past designs, the calculated pressure drop per foot of bed has usually been maintained above 0.25 psi (1.7 kPa) for reaction and 0.2 psi (1.4 kPa) for regeneration, although many reactors such as POWERFORMING units and methanators have operated successfully at lower pressure gradients. For regeneration, a pressure drop of 0.1 psi/ft (2.3 kPa/m) is considered a minimum. These empirical criteria have been found to work for designs in the past. Therefore, it would be desirable to meet these criteria in addition to the minimum Reynolds number requirement presented above. If the design pressure drop is substantially less than these values, the Chemical Engineering Technology Section of EMRE should be consulted.

Fouling Considerations for Reactor Diameter Selection

The typical non-fouling economic design for vapor phase systems will use reactor diameters that produce Reynolds numbers and pressure gradients far above the minimum requirements stated above. For a specific process with a known concentration of foulants (e.g. particulates or diolefins), the rate of foulant deposition in the bed is proportional to the mass flux of the incoming stream. The bed volume available to take up the foulants within specified pressure drop limits is proportional to the bed cross-sectional area. Therefore the typical economic design for fouling systems will require a significantly large reactor diameter to obtain acceptable run lengths.

For SCANfining reactors used in low sulfur mogas applications with FCC naptha type feeds a maximum mass velocity has been set:

total naptha mass velocity < 2500 lb/hr-ft² (w/o catalyst bed bypass)

total naptha mass velocity < 3000 lb/hr-ft² (with catalyst bed bypass)

These limits apply to SCANfining type operations and are not universal across all hydroprocessing units. For other operations these values can be higher or lower depending on the amount of foulants in the bed.

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BASIC DESIGN CONSIDERATIONS (Cont)

To minimize pressure drop through accumulated foulants the hydrodynamic parameter is vapor density times superficial vapor velocity squared (ρv^2). For SCANfining, a second constraint for minimizing pressure drop related problems is:

$$\rho v^2 < 0.4 \text{ (when automatic bypass is not used)}$$

$$\rho v^2 < 0.55 \text{ (when automatic bypass is used)}$$

where: ρ = density of vapor at conditions, lb/ft³

v = superficial vapor velocity ft/sec

These criteria provide the basis for selecting the reactor diameter from fouling considerations. A smaller diameter reactor than specified by these criteria may lead to an unscheduled shutdown due to pressure drop buildup. A diameter larger than specified by these criteria is acceptable provided the minimum Reynolds number and pressure gradient requirements are met.

Inlet Distributors

Three types of inlet distributors are available for use in single-phase packed bed reactors: the slotted distributor, the dish and cone distributor, and the disc and donut distributor (see **Reference 3** and **Figures 1, 2 and 3**). The choice of distributor type depends on the ratio of the effective inlet dynamic head to the total bed pressure drop and on the permissible inlet dynamic head. The purpose of the first criterion is to prevent flow maldistribution in the bed, whereas that of the second one is to prevent movement of the catalyst bed by high-velocity impingement. Criteria for distributor selection are given below.

Slotted Distributor (Figure 1) - The slotted distributor should be used when the effective inlet dynamic head is less than 5% of the total bed pressure drop. The effective reactor inlet velocity to be used for calculating the dynamic head is the actual velocity in the inlet pipe. The allowable effective inlet dynamic heads for various sizes of inert material on top of the bed are listed in **Table 1**.

Dish and Cone Distributor (Figure 2) - Where a small velocity maldistribution in the bed may be harmful to reactor performance, or where the inlet dynamic head is 10% to 30% of the bed pressure drop, a dish and cone distributor is recommended. The upper limits on inlet dynamic head for various sizes of inert material are listed in **Table 1**. If the inlet dynamic head is greater than 30% of the bed pressure drop, the Chemical Engineering Technology Section of EMRE should be consulted.

Disc and Donut Distributor (Figure 3) - The disc and donut distributor should be used when the inlet dynamic head is between 5 and 10% of the bed pressure drop, or when possible high velocity impingement effects on top of the bed are of concern. In grass-roots designs, movement of the inert fill can be minimized by the use of large inert fill. In such cases this distributor offers little advantage, compared to the two designs mentioned above. However, the disc and donut distributor is applicable in special cases, such as unit debottlenecking, when neither the slotted distributor nor the dish and cone distributor is satisfactory because of flow distribution and inert material movement considerations. The allowable inlet dynamic heads for various sizes of inert fills are listed in **Table 1**.

Studies of flow distribution in packed reactors using computational fluid dynamics computer models have shown that the above distribution criteria can be conservative at times. The Chemical Engineering Technology Section of EMRE should be consulted for use of these models if the limitations set by these criteria pose a problem or for situations in which nonstandard vessel geometries (e.g., horizontal drums) or distributors are used.

Quenching

Quenching of single-phase fixed beds should utilize the techniques described in **Section VII-C**. A liquid distributor is not required below the quench hardware device for a vapor quench. When liquid is used as quench in an all vapor phase reactor the hardware must be properly designed to ensure complete vaporization of the quench media prior to contacting the subsequent catalyst bed. Contact the Chemical Engineering Technology Section for vapor phase reactors requiring liquid quench.

Maximum Allowable Pressure Drop

The danger of catalyst crushing sets the maximum pressure drop that can be tolerated during reaction or regeneration, aside from the economic effect on compressor investment and operating cost.

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BASIC DESIGN CONSIDERATIONS (Cont)

Downflow - To avoid the danger of catalyst crushing, the maximum pressure on the catalyst should not exceed 10% of the average single particle crushing strength. The average single particle strength is defined as the crushing force divided by the projected longitudinal cross-section of a catalyst particle. The maximum pressure on the catalyst is calculated by summing the total pressure drop across the bed and the weight per area of catalyst. This criterion is less conservative than the older criterion of 1 psi/ft (23 kPa/m) maximum allowable pressure drop. If needed, a still less conservative criterion is available using data obtained in a bulk catalyst strength tester. For further details on the tester and the correlation to use with the data from the tester, see **Reference 4** or consult the Chemical Engineering Technology Section of EMRE.

Bed Lifting - For downflow regeneration, the maximum pressure drop is not likely to be limiting (see above), and for new designs using extruded catalysts, downflow regeneration should be specified. However, in some past cyclic units using pelleted catalyst, regeneration was carried out upflow. When upflow regeneration is used, the theoretical bed lifting pressure drop must be checked, since bed movement would cause catalyst attrition. See the **DETAILED DESIGN PROCEDURE** Section.

Design Pressure Drop

A safety factor should be applied to the calculated (clean) bed pressure drop during reaction to allow for differences in catalyst loading and for the possible buildup of scale and catalyst fines in the bed. The size and cost of this safety factor will vary with the specific situation. Designs for POWERFORMING units using extruded catalyst have called for 135% of calculated pressure drop, based on the maximum expected catalyst loading in the reactor. In the absence of past experience, a pressure drop of 125% of the calculated clean bed pressure drop should be used. For higher fouling services, such as hydrotreaters fed with cracked or olefinic material, allowances for as much as 100 psi (690 kPa) across the top bed may be required. See **Section VII-A** for a discussion of bed fouling mitigation techniques.

RADIAL FLOW REACTORS (REFERENCE 1)

The general radial reactor configuration is shown in **Figure 4**. The gas flows in a radial direction across the catalyst, which is retained in an annular region by perforated baskets or pipes. Flow may be either radially inward or outward but is usually inward. Radial flow reactors provide high flow surface area per volume of catalyst and thus low bed pressure drop. They are thus desirable when compression costs must be minimized.

Catalyst Contacting Requirements

To assure adequate contacting of the catalyst with the reacting gases, the minimum Reynolds number criterion used in axial reactor designs is applicable. The mass velocity used in the calculation of the Reynolds number should be based on the outer cylindrical surface of the catalyst bed. The minimum Reynolds number is as follows for typical processing reactions:

$$S \quad Re = \frac{W_R D_P}{\mu (1 - \epsilon)} > 100 \quad \text{Eq. (3)}$$

where: Re = Reynolds number, dimensionless
 W_R = Radial mass velocity based on outer surface area of bed, lb/hr ft² (kg/sm²)
 D_P = Equivalent particle diameter, ft (m)
 μ = Fluid viscosity, lb/ft hr (kg/ms)
 ϵ = Void fraction, dimensionless

For regeneration of catalyst a relaxed criterion is acceptable:

$$S \quad Re = \frac{W_R D_P}{\mu (1 - \epsilon)} > 20 \quad \text{Eq. (4)}$$

Flow Direction Selection

There are four possible flow directions that can be used in radial reactors. The gas can flow from the central pipe outward (C-P) or from the periphery inward (P-C). In addition, the flow in the outer annular channel and in the central pipe can be either cocurrent or countercurrent.

For P-C flow, a disc and donut inlet distributor is required to prevent jetting into the annulus. Although the use of an inlet distributor should not present a hardware problem, it does increase the total reactor pressure drop and design complexity.

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BASIC DESIGN CONSIDERATIONS (Cont)

The C-P flow direction can also be subject to inlet jetting and entrance effects if there are abrupt changes in pipe cross-sectional area and in flow direction immediately before the reactor inlet. These cross-sectional area and flow direction changes should be avoided. Flared transition pieces with approximately a 7° half angle spread or turning vanes should be installed when abrupt piping size and flow direction changes are unavoidable.

Cocurrent P-C flow is recommended because of its overall mechanical simplicity. In addition, the P-C flow direction reduces the fouling of the catalyst bed by solids in the feed gas by providing maximum external surface area. It is also recommended that the inlet axial gas flow be in a downward direction.

For processes where severe fouling is expected, the installation of a particle removal device upstream of the reactor should be considered. The bed skimming technique and scale traps used in axial reactors cannot be used in radial reactors. Graded beds are also not practical in radial reactors because they would require the use of additional retaining baskets.

Bed Pressure Drop

The calculation technique for bed pressure drop is given in the **DETAILED DESIGN PROCEDURE** Section.

Flow Distribution Criteria

Adequate gas distribution within the catalyst bed of a radial reactor can be achieved by meeting three flow and geometric criteria. The primary criteria is on β , a parameter related to the ratio of total pressure drop to the axial flow dynamic head in the inlet. The second criteria is on γ which is the ratio of the unpacked annular cross sectional area to that of the center pipe. The third criteria is on the inlet annulus (or pipe) Reynolds number. These parameters were evolved from a mathematical computer simulation of the flows in a radial fixed bed reactor.

The **DETAILED DESIGN PROCEDURE** Section presents the equations used to evaluate these design parameters together with values for these parameters, which result in gas distributions comparable to conventional axial flow fixed bed designs. To reduce the built in conservatism in the design, with possible savings in investment and operating costs, a computer model should be utilized. This model is available as Program No. D3695 (**Reference 5**). The program, when run with a criteria of 4% maldistribution will give gas distribution equivalent to an axial flow fixed bed reactor utilizing a slotted or disc and donut distributor. Using a 1% maldistribution criteria will give a design equivalent to conventional axial flow fixed beds using a dish and cone distributor.

Catalyst Bed Sizing

There is no limit on the vessel size or reactor height to bed thickness ratio. It is necessary, however, to set a minimum bed thickness to ensure plug flow behavior and prevent gas bypassing caused by inlet jetting. This minimum bed thickness is 50 equivalent particle diameters:

$$r_1 - r \geq 50 D_p \quad \text{Eq. (5)}$$

where: r_1 = Radius of outer catalyst bed, ft (m)
 r = Radius of inner catalyst bed, ft (m)

Catalyst Basket Design

Several types of construction are possible for catalyst baskets: Johnson Profile screen, wedge wire screen, scallop screens and wire wrapped perforated pipe. There is no fluid dynamic advantage for any of these alternatives. These materials are all acceptable for reactor design. The choice should be made on the basis of mechanical and economic considerations.

Criteria to be used in the design of catalyst baskets with screens or perforated pipe are given in the **DETAILED DESIGN PROCEDURE** Section.