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Revision Memo

08/04 This is an error correction to Bed Dimensions Section. Change made 4.1 kg/s/m² to 2.7 kg/s/M²



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SCOPE

This section gives recommended general process and mechanical design procedures for packed bed reactors, including fixed bed, radial, and tubular reactors and for desiccant driers to be used for single-phase (all-vapor or all-liquid) or mixed-phase (liquid and vapor) service. If a single-phase reactor is being designed, this section should be used together with **Section VII-B**, while for mixed-phase designs it should be used along with **Section VII-C**.

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. (51 mm)
- GP 5-2-2 Internals for Fixed Bed Reactors

OTHER LITERATURE

1. *Design Guidelines for Fixed Bed Vapor Phase Inlet Distributors*, ER&E Report No. EE.3E.74, January, 1974.
2. *Design Guidelines for Maximum Allowable Pressure Drop in Fixed Beds*, ER&E Report No. EE.11E.74, December 1974.
3. *Techniques for Mitigating Fouling of Mixed Phase Downflow Fixed Bed Reactors*, ER&E Report No. EE.42E.81, May 1981.
4. *Guidelines for Mitigating Fouling, Plugging, and Corrosion in Hydroprocessing Units*, ER&E Report No. EE.107E.96, October 1996.
5. *Diagnostic and Troubleshooting Tools for Fixed Bed Reactors*, ER&E Report No. EE.137E.91, December, 1991.

BACKGROUND

Fixed bed reactors are used in numerous ExxonMobil petroleum and chemicals processes. They consist of a cylindrical or spherical vessel in which one or several beds of catalyst particles are located. A single or two-phase flow enters the vessel and contacts the catalyst. In an axial flow reactor, flow is either downward or upward. Most applications with two-phase flow are co-current down flowing reactors. In a few applications counter-current flow with liquid downflow and vapor upflow is preferred. In a radial flow reactor, flow either enters a central perforated pipe or basket and flows radially outward, or it enters an outer annular region and flows radially inward. The latter mode is more typical.

Fixed beds except for cyclic **POWERFORMING** units, are usually used in processes in which catalyst deactivation takes place on a scale of months or years. They can be used for either exothermic or endothermic reactions. Highly exothermic reactions can be carried out in tubular reactors where catalyst is packed in tubes with diameters no greater than several inches. This limit on tube diameter is set by the need to provide large surface area to transfer reaction heat through the tube wall. The term "fixed bed" is normally reserved for beds with diameters on the order of a ft or more. Heat removal through the wall of a fixed bed is not practical due to the low surface to volume ratio and poor radial heat transfer. Fixed beds are advantageous when compared to fluidized beds because of their ability to handle a given throughput in a smaller diameter vessel and because of their simplicity. Radial flow reactors provide a high surface area for flow per volume of catalyst and thus give low-pressure drop. For some cases reaction rate inhibitors can be removed in counter-current operation decreasing the required catalyst volume.

Important considerations which must be addressed in the design of fixed bed reactors include: bed pressure drop, catalyst loading and unloading, distribution of process fluids to the bed, potential to foul the bed by physical or chemical foulants, temperature monitoring and the possibility of generating localized hot spots. For counter-current reactors flooding is an important concern. For information on flooding and design issues contact the Chemical Engineering Technology Section of EMRE. These items will be addressed in detail in the sections below.



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DEFINITIONS

BED REYNOLDS NUMBER

The bed Reynolds number expresses the ratio of kinetic to viscous forces in a flow field within the packed bed.

$$Re_b = \frac{W D_p}{\mu (1 - \varepsilon)} \quad \text{Eq. (1)}$$

where: Re_b = Bed Reynolds number, dimensionless
 W = Mass velocity (gas or liquid) lb/hr ft² (kg/sm²) based on reactor cross-sectional area
 D_p = Equivalent particle diameter, ft (m)
 μ = Viscosity at conditions, lb/ft hr (= centipoises x 2.42) (kg/ms)
 ε = Void fraction of bed, dimensionless

EQUIVALENT PARTICLE DIAMETER

The equivalent particle diameter is the diameter of a sphere which has the same volume to surface area ratio as that of the catalyst particle.

$$D_p = \frac{6 V_p}{S_p} \quad \text{Eq. (2)}$$

where: D_p = Equivalent particle diameter, ft (m)
 V_p = Volume of a catalyst particle, ft³ (m³)
 S_p = Surface area of the catalyst particle, ft² (m²)

FIXED BED REACTOR

A vessel with diameter of about 1 ft or larger packed with catalyst particles through which a single or two-phase mixture of fluids flows and reacts.

RADIAL FLOW REACTOR

A vessel in which catalyst particles are contained in a cylindrical annular region. A gas phase mixture flows radially inward or outward and reacts.

REGENERABLE DESICCANT DRIERS

Regenerable desiccant driers are vessels containing packed beds of non-consumable desiccant, which remove water from a liquid or vapor stream by adsorption. The beds are periodically regenerated in cyclic operation. Examples of such desiccants include molecular sieves, activated alumina, and silica gel. Specifically excluded are salt driers and sand beds, which operate on liquid/liquid streams by a combination of coalescence and adsorption.

TUBULAR REACTOR

A cylindrical pipe with diameter less than about 1 ft packed with catalyst particles. A single or two-phase mixture of fluids flows through the packing and reacts.

BASIC DESIGN CONSIDERATIONS

MECHANICAL AND ECONOMIC CONSIDERATIONS

The optimum cost reactor should be determined based on process requirements, together with mechanical and materials considerations. The effect of reactor diameter and height and the number of reactors on total reactor cost must be balanced against investment and operating costs for the treat gas compressor (when required), pumps, external piping, regeneration facilities, etc. Reactor type (lined, unlined, alloy clad, hot or cold wall) and shape (cylindrical, spherical) are also factors to be considered. See **Section VII-D (Mechanical Details)** for selection and mechanical design considerations regarding reactor type and shape and for shrouds for protection against corrosion and hydrogen attack.

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BASIC DESIGN CONSIDERATIONS (Cont)

REACTOR VOLUME

In most cases the volume occupied by the catalyst is specified in the DBS. (For some units such as **POWERFORMING** units, the DBS specifies the total weight of catalyst. The catalyst volume is determined from the total weight of catalyst and catalyst bulk density.) In addition to the volume occupied by the catalyst, room must be provided for reactor internals, inert fill above and below the catalyst bed, spare volume for extra catalyst (if specified in the DBS), nozzle clearance, and access for inspection.

CATALYST SPECIFICATIONS - PARTICLE DIAMETER AND VOID FRACTION

The shaped catalyst with three or four lobes has replaced the cylindrical extrudate as the catalyst of choice for fixed bed applications. The main ones in use are the asymmetric quadralobe (ASQ) and trilobe. Symmetric quadralobe catalysts are also used in some applications. The asymmetric quadralobe shape, shown in **Figure 1**, allows for the use of smaller diameter particles and thus higher activity without the higher-pressure drop normally associated with smaller catalyst size.

The equivalent particle diameter to be used in the bed pressure drop calculations (see **Sections VII-B** and **VII-C**) is given for cylindrical particles as:

$$D_p = \frac{3DL}{2L + D} \quad \text{Eq. (3)}$$

where: D_p = Equivalent particle diameter, ft (m)
 D = Actual particle diameter, ft (m)
 L = Average particle length, ft (m)

For 1/20 in. (1.3 mm) asymmetric quadralobe the equivalent diameter is given by the specific formula shown in **Table 1**. For the ASQ catalyst in general the procedure shown in **Table 2**, should be used to calculate the equivalent particle diameter.

The bulk density of commercial catalyst beds can vary over wide limits, depending on the shape and particle size distribution of the catalyst and on the techniques used for loading the catalyst into the reactor. Some catalyst beds may settle with time, thereby increasing the catalyst bulk density. Commercial data on this effect are inconclusive, sometimes indicating bed settling and sometimes not. Values for expected commercial bed packing and void fraction for freshly packed beds are given in **Table 1**. The nominal values in **Table 1** should be used for calculating required catalyst volume or weight from the catalyst weight or volume specified in the DBS unless specific data for the actual catalyst in use is available. However, because of the wide range of possible bed void fractions, it is recommended that a bed void fraction (ϵ) of 0.35 for cylindrical extrudates and 1/8 in. (3.2 mm) pills and 0.4 for ASQ and trilobe catalyst be used for clean bed pressure drop calculations which form the basis for compressor sizing.

For data on catalyst shapes and sizes not shown in **Table 1**, consult the Chemical Engineering Technology Section of EMRE.

BED DIMENSIONS

When the catalyst volume has been set, the bed diameter and height must be determined. Plant experience has shown that the bed geometry can vary over a reasonably wide range and still satisfy process requirements. Process restrictions which do apply concern fluid velocities and bed pressure drops. The lower limit on bed height is based on achieving good distribution across the bed and good contacting throughout the bed, whereas the upper limit is based on preventing catalyst crushing and the creation of fines and excessive temperature buildup or reduction for exothermic and endothermic reactions respectively. For single-phase reactors fluid is quickly distributed by the packing within a short distance from the bed entrance. In addition, axial dispersion is low and plug flow is easily achieved with a shallow bed. A practical limit for minimum bed depth is 1 ft (305 mm) below the bottom of scale traps if present. In two-phase trickle flow, however, initial maldistributions can propagate for significant distances into the packing if liquid mass velocity is not high enough. Axial liquid dispersion in trickle flow is considerably higher than in single-phase flow. Therefore to maintain plugflow, trickle flow beds should be no shorter than 4 ft (1219 mm). If liquid mass velocity is below 2000 lb/hr ft² (2.7 kg/sm²), and bed length is less than 10 ft (3048 mm), the Chemical Engineering Technology Section of EMRE should be consulted to determine liquid distribution and contacting within the bed for bubble cap and tube and chimney distributor trays (see **Section VII-C**). The spray nozzle distributor and the downcomer spray trays provide adequate distribution to the top of the bed over the full range of liquid mass velocities.

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BASIC DESIGN CONSIDERATIONS (Cont)

The main economic factor is the effect of reactor geometry on the size of the recycle gas compressor. A decrease in vessel diameter (increasing the length) increases bed pressure drop, which in turn increases compressor investment and operating costs for a given recycle gas rate. Reactor shell costs, particularly for heavy-wall vessels, decrease with decreasing diameter. Thus, recycle gas compressor investment and operating costs must be balanced against reactor shell costs. The optimum length to diameter ratio for a given bed depends upon the size of the unit, the specific location, catalyst type, regeneration and treat gas rates, and for POWERFORMING units, the type of unit (semiregenerative or cyclic). The need for liquid redistribution in a two-phase unit is discussed in **Section VII-C**.

STANDARD SHELL SIZES

For economic reasons, the bed diameter of cylindrical reactors should be adjusted to the nearest dimension, which will permit the use of standard reactor heads, and the bed pressure drop should be recalculated for the new diameter. In inch sizes, standard elliptical heads in shell thickness normally used for catalytic reactors are available in 6 in. diameter increments for *inside* diameters from 3 to 18 ft. In metric sizes, standard elliptical heads are available in 100 mm increments for diameters from 300 mm to 3000 mm, and in 200 mm increments for diameters from 3000 to 3600 mm. The standard metric diameter is measured *outside* in Europe and *inside* in the Far East. In relating *bed* diameter to *shell* diameter, one must take into account such items as the lining, shroud thickness and shroud clearance, and for metric sizes, shell thickness.

FLUID FLOW AND CONTACTING

Efficient utilization of the catalyst in a packed bed reactor requires good contacting between the vapor and/or liquid reactants and the catalyst particles. General considerations that affect catalyst contacting efficiency are discussed in this section. Specific design criteria for vapor phase or mixed-phase fixed bed reactors are covered in **Sections VII-B** and **VII-C**.

Good catalyst performance depends on uniform distribution of the reactants to the bed inlet and on the proper fluid dynamics (velocity and flow uniformity) within the catalyst bed. Uniform flow distribution to the bed inlet is determined by the design of the inlet distributor and by the relationship between the kinetic energy of the inlet stream and bed pressure drop. To ensure uniform flow through the catalyst bed and effective contacting between reactant streams and the catalyst, minimum velocity criteria must be met. For all-vapor or all-liquid phase packed bed reactors, these velocity criteria are normally given in terms of minimum bed Reynolds numbers. For SCANFINING units, additional criteria have been developed to insure high selectivity to sulfur removal. Contact the Chemical Engineering Technology section of EMRE for further details. For mixed-phase packed bed reactors, the velocity criteria are stated in terms of liquid mass velocity and pressure drop requirements.

In regenerative units, fluid compositions and mass flow rates during reaction differ markedly from those during regeneration. The *maximum* allowable velocity and pressure drop are usually set by reaction operation, whereas the *minimum* allowable velocity and pressure drop may be set by either reaction or regeneration. Both operations should be checked for pressure drop, and for uniform flow and contacting.

In some of the discussions in this section, the bed inlet is referred to as the "top" of the bed and the outlet as the "bottom" of the bed. This terminology simply reflects the fact that downflow reactors outnumber upflow reactors by a wide margin, and is not intended to suggest that all reactors *must* be downflow.

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.

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 unit area (in consistent units) of the catalyst and of liquid holdup if there is liquid in the feed. 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 1** or consult the Chemical Engineering Technology Section of EMRE.

TEMPERATURE INSTRUMENTATION

Reaction and Regeneration - Temperature control is extremely important with highly exothermic reactions (which normally require the staged addition of cold quench gas or liquid), and during catalyst activation and regeneration. In such situations, temperature runaways can cause permanent catalyst deactivation and safety issues. These reactors require close thermocouple spacing to assure good temperature monitoring and/or control. In reactors where close temperature control is less important, wider thermocouple spacing can be used.

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BASIC DESIGN CONSIDERATIONS (Cont)

There is no general algorithm for the prescription of the type and number of thermocouples which should be used in the design of fixed, radial, or tubular reactors. The primary consideration must be the adequate monitoring and control of reactor temperatures to protect the vessel against overheating. The exothermicity of the reactions involved, bed length and diameter, flow distribution, bed fouling tendency, and required thermocouple response time will guide in the selection of the temperature instrumentation. For highly exothermic reactions a reactive chemistry laboratory study and mathematical modeling effort should be considered to help define needed instrumentation. These studies define the size and the temperature of potential hot spots based on kinetics, bed hydrodynamics and heat transfer properties. The proper thermocouple spacing can then be selected to ensure detection before the bed reaches a specified temperature or the vessel design temperature. For reactors operating close to the vessel design temperature external, skin thermocouples may be required to ensure that the vessel design temperature is not exceeded.

THERMOCOUPLE TYPES AND LAYOUT

There are several different types and layouts which have been used for temperature control and monitoring in fixed bed and radial reactors. (See **Section VII-B** for instrumentation for tubular reactors.) These include:

- Top or bottom entering thermowells with multiple thermocouples spaced vertically along the well. These are used mainly for monitoring since their response is slow. These thermowells are adequate when fluid maldistribution and fouling are not anticipated. They are positioned on a circle whose area equals one-half the cross-sectional area of the bed. If the bed diameter is less than 3 ft (914 mm), two thermowells 180 degrees apart are adequate. For larger reactors, three wells should be provided 120 degrees apart. Four or more wells can be used for beds of 12 ft (3658 mm) diameter or larger. Vertical spacing is determined by anticipated axial temperature rise. In the past, the temperature points have been staggered through the bed if two wells are used, or arranged in a spiral pattern if three are used (see **Figure 2**). Recently, for Hydrocrackers and other exothermic units, the thermocouples within the wells have been placed at the same level to simplify assessment of possible temperature and flow maldistributions in the reactor.
- Side entering thermowells with multiple thermocouples along the well. These have been used in Hydrocrackers. They can be high speed and can be used for control or they can be slower response thermowells as used in vertical applications (see **Figure 3**). These thermowells may be used to monitor axial temperature profiles by stacking a series of them in a vertical plane. They may also be used to assess temperature maldistribution in a horizontal cross-section by having multiple wells enter the vessel at the same level but at different azimuthal locations.
- Flexible thermocouples individually placed or in radial assemblies. These are the preferred method for monitoring temperatures in a cross-section of the reactor and should be used for highly exothermic systems (Hydrocrackers and Hydrogenation units) where maldistribution due to fouling is a possibility. **Figure 4** shows an example of this assembly. The Mechanical and Instrumentation Sections of EMRE should be consulted for construction details.
- Skin thermocouples attached to the outer wall of the reactor toward the bottom of each bed and on the bottom head. These generally have slow response and are biased toward temperature readings below the inside temperatures due to heat losses. The thermocouples should be considered for highly exothermic systems such as Hydrocrackers and aromatics hydrogenation units to give added monitoring capability for detecting high temperature regions. There are two types of skin thermocouples:
 - Individual thermocouples welded to the reactor wall at intervals of about 2 ft (610 mm).
 - Continuous thermocouple line detectors. These do not pinpoint the hot spot but can indicate that a hot spot exists anywhere along its length.

There are no set guidelines for the number of internal and/or skin thermocouples which should be used. The appropriate process section and the Chemical Engineering Technology Section of EMRE should be consulted for guidance.

INERT FILL

In addition to the catalyst, the reactor contains inert fill at the top to prevent catalyst movement and to help distribute the inlet flow; at the bottom, to support the catalyst beds; and at times in the void volume which results if the reactor is initially loaded with less than a full charge of catalyst. Inert fill can also be used to separate layers of different catalyst.

The fill must be chemically inert, thermally stable, and free of catalyst poisons. Typically, inert fill material is pure Al_2O_3 or mixtures of Al_2O_3 and SiO_2 , usually in the form of spheres or pellets. Pelleted inert fill, which can be obtained in sizes between 1/8 in. x 1/8 in. (3.2 mm x 3.2 mm) to 7/8 in. x 1-1/4 in. (22.2 mm x 31.8 mm), often has a substantial price advantage over the same base material in spherical form. However, pelleted fill is more susceptible to attrition because it has edges.