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Revision Memo

- 12/01** This update consists primarily of an incorporation of heritage Mobil technology for pressure drop prediction, distributor tray design and the spider vortex quench assembly design.



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SCOPE

This section gives recommended process and mechanical design procedures for mixed-phase (liquid and vapor) packed-bed reactors. Flow through the bed is cocurrent and can be either upward or downward. General design procedures applicable to all types of reactors are given in [Section VII-A](#). Detailed mechanical considerations, shrouds, linings, and corrosion problems are covered in [Section VII-D](#).

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

- Larkins, R. P., White, R. R., D. W. Jeffrey, AIChE J, 7, 231 (1961).
- Weekman, V. M., Meyers, J. E., AIChE J, 10, 951 (1964).
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- Update on Pressure Drop Estimation in Mixed Phase Fixed Bed Reactors*, ER&E Report No. EE.38.E.99, December, 1998.
- An Improved Method for Predicting Two-Phase Pressure Drop in Packed Beds*, MTC Report No.96M-0654, December, 1996.
- Implementation of the Parallel Flow Method for Prediction of Two-Phase Pressure Drop in Fixed Beds*, MTC Report No. 97M-1152, December 1997.
- Quantitative Studies of Gas-Liquid Flow in Bubble Caps and Downcomers Using a High-Accuracy Cold Flow Apparatus*, MTC Report No.DAN 92M-689.
- Experimental Evaluation of Fixed Bed Reactor Inlet Nozzle Distributors*, MTC Report No.DAN 98-M-0358, April, 1998.
- Fixed Bed Reactor Quench Assembly Design Guidelines*, ER&E Report No. EE.82E.91, December, 1991.
- Techniques for Mitigating Fouling of Mixed Phase Downflow Fixed Bed Reactors*, ER&E Report No. EE.42E.81, May 1981.

DEFINITIONS

See [Section VII-A](#).

BACKGROUND

While the direction of flow of the vapor-liquid feed may be either upward or downward, historically, downflow reactors have been used almost exclusively to prevent problems associated with catalyst fluidization. Downflow operation should be used for all designs in which it is expected that complete vaporization of the feed will occur in the bed (feed passing through the "dry point"). Most of the design procedures covered in this section pertain only to downflow operation. For upflow design considerations the Chemical Engineering Technology Section of EMRE should be consulted.

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APPLICATIONS

The following processes or services are examples of applications in which mixed-phase, packed-bed reactors are used:

- Hydrocrackers
- Hydrodenitrogenation units
- Hydrofiners (lube & wax, gas oil, cat feed, steam cracked naphtha)
- Gofiners (vacuum gas hydrodesulfurization)
- Residfiners (residuum hydrodesulfurization)
- Liquid phase hydrogenation units (e.g., oxo alcohol finishing)

BASIC DESIGN CONSIDERATIONS

FLUID FLOW AND CONTACTING

In downflow mixed-phase packed-bed reactors, liquid-catalyst contacting is usually the controlling factor for effective catalyst utilization. If uniform liquid distribution and good contacting between liquid and catalyst are obtained, vapor-liquid contacting will generally not be a problem.

In terms of contacting, the most critical mixed-phase services are those in which high conversion (e.g., greater than 80%) is desired. To achieve these conversions, it is usually necessary to operate the reactor at low space velocity. This includes most lube and wax hydrofiners and some desulfurization reactors. Therefore, the following criteria should be tempered by consideration of the severity of the operation in question.

Good contacting efficiency and catalyst utilization require: (a) good liquid distribution to the top of the catalyst bed, (b) uniform flow of the reactants through the catalyst bed, (c) sufficient pressure drop through the bed, and (d) high liquid mass velocities. These factors can be influenced by the catalyst size and shape, the manner in which the catalyst is packed, and the bed startup procedure. Catalyst should be loaded into the reactor in small increments and raked level across the bed to insure a uniform, dense packing. Sock loading of catalyst with frequent raking or automatic catalyst dense loading methods are both acceptable.

High liquid loading (mass velocity) is necessary for good catalyst utilization. Where feed vaporization is slight (less than 20 to 30 wt %) liquid loading (based on actual liquid rate in the reactor at conditions) should be maintained above 3000 lb/hr ft² (4.07 kg/sm²) at design feed rates. Above this loading it may be assumed that 100% catalyst utilization is achieved. If the reactor must be designed for lower liquid mass velocities than this, a contacting debit is incurred and the catalyst volume should be increased by 10% (assuming 90% contacting efficiency) to compensate for poorer contacting. Mass velocities less than 2000 lb/hr ft² (2.72 kg/sm²) at the bed inlet should not be used for design feed rates.

In cases where significant feed vaporization occurs (greater than 20 to 30%), it may be difficult to design for a liquid loading of 3000 lb/hr ft² (4.07 kg/sm²) at design conditions without encountering high bed pressure drops (> 1 psi/ft) (23 kPa/m). In these situations, the unit should be designed for as high an inlet mass velocity as possible, consistent with bed pressure drop constraints and limitations of available treat gas pressure. Normally, a contacting efficiency debit need not be applied in these cases, but the lower limit on mass velocity of 1200 lb/hr ft² (1.63 kg/sm²) anywhere in the bed should not be violated, unless the reactants go through the dry point in the reactor.

In designs with high turndown ratios, it is often impossible to maintain the minimum liquid mass velocity above 3000 lb/hr ft² (4.07 kg/sm²) in all operations. In these cases extra catalyst is normally available at the lower throughputs, i.e., space velocity is lower than design, so no contacting efficiency debit need be applied. As before, liquid mass velocity should be maintained above 1200 lb/hr ft² (1.63 kg/sm²) at all points in the bed.

S LIQUID DISTRIBUTORS

Uniform initial liquid distribution is important for achieving good catalyst utilization. Leakage rates of installed trays must be less than 1% of design flow at conditions or about 3 gph/ft² (3.4 x 10⁻⁵ m³/s/m²) for a mass velocity of 3000 lb/hr ft² (4.07 kg/sm²). The tray must also be level to within 1/900th of the tray diameter (as specified in Global Practice **GP 5-2-2**) to avoid significant losses in catalyst utilization. Five types of distributors are currently in use in ExxonMobil mixed-phase, packed-bed reactors. They are the spray nozzle tray, downcomer spray tray, short tube-slotted chimney tray, perforated plate, and bubble cap tray. The following criteria should be used for selecting the appropriate distributor:

Spray Nozzle Tray - The spray nozzle tray is shown in **Figure 1**. Overlapping spray from spray nozzles produces the most uniform liquid distribution to the top of catalyst beds at all liquid mass velocities. However, spray trays cost up to twice as much as tube and chimney designs. This type of tray is recommended for liquid mass velocity below 2000 - 2500 lb/hr ft² (2.71 - 3.39 kg/sm²) for beds less than 10 ft (3048 mm) and liquid mass velocity below 1500 - 2000 lb/hr ft² (2.03 - 2.71 kg/sm²) for beds longer than 10 ft (3048 mm). This tray should also be used under fouling conditions where a significant number of tubes plug at any liquid rate. It should also be used following injection of quench gas to achieve better mixing of the vapor and liquid.

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BASIC DESIGN CONSIDERATIONS (Cont)

- Chimney Downcomer SprayTray** - The downcomer spray tray is shown in **Figure 2**. In this design liquid jets formed from flow through holes on the side of the tubes is sheared into droplets by the vapor coming down the tube. The overlapping spray from downcomer tubes produce a uniform liquid distribution similar to that produced from the spray nozzle tray. This type of tray is always used with a calming tray above it. The downcomer spray tray can be used in place of spray nozzle trays except for conditions where significant fouling may occur and plug the holes.

Short Tube-Slotted Chimney Tray - (**Figure 3**) - Under normal conditions of high liquid mass velocity the short tube-slotted chimney distributor should be specified. This type of distributor is effective for liquid mass velocities above 1500 - 2000 lb/hr ft² (2.03 - 2.71 kg/sm²) for long beds and 2000 - 2500 lb/hr ft² (2.71 - 3.39 kg/sm²) for short beds.

Perforated Plate - It is recommended that all existing perforated plate distributors be replaced. **Bubble Cap Tray** - () - The bubble cap distributor has been used in reactors in which quench gas is injected between beds and where a significant portion of the heat transfer is effected by the distributor. It has been replaced by the spray nozzle tray and is no longer recommended for new designs. However, because many existing reactors have this type of distributor, design procedures are included in this section.

S LIQUID REDISTRIBUTION

To reduce possible maldistribution, liquid redistribution is normally used in deep catalyst beds. The same trays used for initial liquid distribution in the reactor are used for redistribution. Liquid redistribution should be considered when conversion in the reactor is greater than 85% or when bed fouling is expected. It is recommended that the reactor be split into at least two beds for conversions between 85 and 95% and at least into three beds for conversions exceeding 95%. For exothermic reactions, bed height needs to be limited based on the controllability of each bed and the potential for undesired reactive chemistry outside the normal operating temperature range. These undesired reactions can occur due to accidental malfunctions of internals or due to process trips or accidental maloperation. It is recommended that for all exothermic reactions, the reactive chemistry issues be reviewed and understood before setting maximum bed heights. The Chemical Engineering Technology Section of EMRE should be consulted for advice on these matters.

For recommendations on bed splitting for fouling feeds see the section on FIXED BED FOULING.

QUENCHING FLUID AND QUENCH NOZZLES

Intermediate cooling of reactor contents is often required in highly exothermic systems such as hydrocracking. This is normally accomplished by injecting cold fluid between catalyst beds. The cold fluid can be either a gas or a gas / liquid mixture. An important design consideration for interbed quench assemblies is that they completely mix the quench with the process fluids as well as uniformly distribute flow to the next catalyst bed. This will reduce the possibility of generating hot spots in the bed.

This section describes the design of the swirl cap fixed bed reactor quench assembly. The guidelines cover the sizing of the quench injection sparger, swirl mixing cap, and annular donut baffle. Programs, which perform the design calculations, are available (CPEE090). The improved quench assembly design described in this section is recommended for all grass roots fixed bed reactors requiring interbed quench. Significant incentives may exist for replacing the older quench assemblies with this improved design. Although the older quench box design is no longer recommended, a design procedure is included in this section to evaluate existing designs.

A description of the spider vortex quench assembly is included below. For detailed design guidelines and procedures contact , the Chemical Engineering Technology Section of EMRE.

An alternative quenching technique, which has sometimes been used (with small diameter reactors), is to introduce gas quench in the transfer line between reactors in series. This method eliminates the need for gas quench hardware and, where appropriate, may be more economical. Generally, twenty pipe diameters after injection into a turbulent, bubbly, or churn flow transfer line will assure adequate mixing. In cases where these conditions do not obtain, the Chemical Engineering Technology Section of EMRE should be consulted.

MINIMUM PRESSURE DROP

If calculated pressure drop through the bed is less than 0.1 psi/ft (2.3 kPa/m) for either reaction or regeneration, Chemical Engineering Technology Section of EMRE should be consulted.

MAXIMUM ALLOWABLE PRESSURE DROP

Pressure drop in mixed-phase packed-bed reactors has ranged principally from below 0.1 psi/ft (2.3 kPa/m) to 1.0 psi/ft (23 kPa/m). 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 unit area (in consistent units) of catalyst, and 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 5** or consult the Chemical Engineering Technology Section of EMRE.

Bed Lifting - For downflow regeneration, the maximum pressure drop is not likely to be limiting, and for new designs using extruded catalyst, downflow regeneration should be specified. For any upflow operation, the theoretical bed lifting pressure drop must be checked, since bed movement would cause catalyst attrition. See BED LIFTING below.

DESIGN PRESSURE DROP

For the purpose of setting system pressure balance or recycle gas compressor requirements a safety factor should be applied to the clean bed pressure drop calculated for reaction conditions to allow for differences in catalyst loading and for the possible buildup of coke, scale and catalyst fines in the bed. The design pressure drop for new units should be twice the calculated clean bed pressure drop or the calculated clean bed pressure drop plus 50 psi (345 kPa) whichever is less. For revamped units, the pressure drop allowance for fouling should be based on plant data if possible.

For designing *bed supports* in multiple-bed reactors, the following additional fouling allowances should be added to the sum of the maximum operating pressure drop under fouled conditions and the weight per unit area of catalyst, inert fill and liquid holdup. The weight of catalyst should be calculated based on a bulk density of 60 lb/ft³ (962 kg/m³), regardless of the actual catalyst being used.

Beds with Scale Traps (Severe fouling service) - Add weight equivalent to 40 psi (276 kPa) fouling allowance for each bed.

Beds without Scale Traps (Low fouling service) - Add weight equivalent to 10 psi (69 kPa) fouling allowance for each bed.

See **Section VII-A** for a discussion of the use of scale traps.

BED LIFTING

The possibility of bed lifting must be considered whenever mixed-phase reactors are operated upflow. This will occur whenever the pressure drop across the bed equals the effective weight per area of the bed. Since generalized methods for predicting gas holdup and bed pressure drop in upflow, mixed-phase packed-bed reactors are not yet available, it is suggested that Chemical Engineering Technology Section of EMRE be consulted whenever this situation is encountered.

CATALYST DUMP TUBES

The designs of several recent multiple bed reactors have included catalyst dump tubes between beds. The purpose of these tubes is to allow all of the catalyst in the reactor to be removed through the bottom catalyst unloading nozzle. It is recommended that when catalyst dump tubes are used, a side catalyst unloading nozzle should also be specified at the bottom of each bed above the bottom bed to enable the option for rapid and effective dumping of the catalyst from the reactor. Alternatively, in difficult fouling situations, provision for vacuuming out the catalyst would have to be considered. Care should be taken that the reactor internals do not interfere with the catalyst unloading nozzle.

FIXED BED FOULING

A general discussion of the important considerations relating to fouling of mixed phased fixed beds is presented in **Section VII-A**. Design guidelines for various techniques used to mitigate fouling are given in the **DETAILED DESIGN PROCEDURE** Section. These include backwashing filters, sand filters, scale traps, guard reactors, grading, feed blanketing, and bed bypassing. This last technique is now discussed further.

Fouling tolerance of fixed bed hydroprocessing reactors can be increased and run lengths extended by using automatic fouled catalyst bypass technology. This technology has been commercially demonstrated in several units processing mid-distillates, vacuum gas oils and resids. The bypass technology may be applied to any hydroprocessing reactor with two-phase gas-liquid flow where the catalyst bed top is susceptible to fouling due to deposition of coke, metals, polymers, sludge and feed contaminant particulates. For designs involving highly exothermic reactions (such as hydrocrackers) the Chemical Engineering Technology Section of EMRE should be consulted prior to use of this technology.

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BASIC DESIGN CONSIDERATIONS (Cont)

The fouled catalyst bypass design involves providing a short catalyst bed with bypass pipes above the main catalyst bed(s) as is shown in **Figure 4**. Fouling slowly increases the pressure drop across this short bed. This forces gradually increasing fractions of gas and liquid to bypass the bed. Without bypassing, a foulant crust of low permeability would form which becomes a filter for very small contaminant particles in the feed. This leads to rapid pressure drop buildup and eventual reactor shutdown. The bypass achieves moderation in pressure drop buildup by exploiting several synergies between fouling and two-phase fluid dynamics:

- Spreads foulants over two beds thereby preventing/delaying fouling crust formation.
- Reduces the net flow through fouled/crusted catalyst layer thereby reducing the reactor pressure drop.
- Leads to a gradually increasing segregation of liquid and gas flows since separate bypass pipes are provided for the liquid and the gas. Since the single phase pressure drops are an order of magnitude less than mixed-phase, gas-liquid pressure drop, very significant further reduction in pressure drop is achieved.
- Permits the top bed to continue filtering a fraction of the feed even after crusting because some bed permeability is always retained. Thus, the top fouled bed is never abandoned and continues to play a role in minimizing lower bed fouling.

Experience has shown that the bypass technology can increase run length by a factor of four in fouling constrained reactors. Further improvements may be accomplished by using multiple tiers of bypass beds. The bypass technology saves the investment of installing a separate guard reactor. Also, it eliminates the cost of periodic guard reactor changeouts. With the run length increase, many hydroprocessing units will no longer be pressure drop constrained but only constrained by catalyst deactivation.

DETAILED DESIGN PROCEDURE

MECHANICAL CONSIDERATIONS

See **Sections VII-D** and **VII-E** for mechanical design procedures.

♦ BED PRESSURE DROP

Significant efforts have been made in the past five years to revise and improve pressure drop prediction in two-phase packed beds. The heritage Exxon 1995 Design Practice correlations were reviewed using a compilation of commercial and cold flow data covering a wide range of catalyst shapes, sizes, and feed properties (**Reference 6**). It was concluded that the DP correlations for both high and low boiling range feeds provide estimates which are consistent with commercial pressure drop measurements and within the band of uncertainty that should be expected from variations in void fraction and particle size.

During this same time period a separate extensive study by heritage Mobil (**References 7 and 8**) resulted in an improved method for pressure drop prediction in packed beds. The parallel flow method is a more fundamental model used to predict two-phase pressure drop in packed beds. The technique has two key features. First, the packed bed is described in terms of flowing void and pressure-neutral void, the latter being interparticle channels through which negligible flow occurs because the pressures at both ends are nearly equal. Pressure neutral void is unique to a given catalyst size, shape, and loading and is determined by a gas flow experiment. Second, gas-liquid passage through the flowing void is represented as parallel flows described by Ergun equation for each phase—there are no case-specific or other fitted parameters. Liquid holdup is determined by the constraint that the pressure drops in each phase be equal across the catalyst.

Comparison of the two prediction methods have been made for only a few units and a comprehensive study has not been completed. Both methods are therefore presented here. For new designs, calculations using both methods are recommended. For analysis of apparent contradictory results contact the Chemical Engineering Technology Section of EMRE.

Heritage Exxon Old Design Practice Correlation

Feeds with Final Boiling Points Below 800°F (700°K)

The procedure given below and the correlation presented in **Figures 6 and 7** should be used for calculating two-phase pressure drop for co-current downflow operation with liquids with final boiling points below 800°F (700°K). The correlation below is based upon a method developed by Larkins, White and Jeffrey (**Reference 1**) and by Weekman and Meyers (**Reference 2**) and extended by Exxon Engineering (**Reference 3**). The following equations must be used, to the exclusion of other possible single-phase pressure drop equations, since the correlations used in this procedure were developed based on **Eqs. (1) and (2)**.

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BASIC DESIGN CONSIDERATIONS (Cont)

Step 1: Gas Phase Pressure Drop - Calculate $(\Delta P/L)_G$, the pressure drop which the gas phase would develop if it were present alone, from the Ergun equation:

$$\left(\frac{\Delta P}{L}\right)_G = K \operatorname{Re}_G (150 + 1.75 \operatorname{Re}_G) \left(\frac{1-\varepsilon}{\varepsilon}\right)^3 \left(\frac{\mu_G^2}{\rho_G D_p^3}\right) \quad \text{Eq. (1)}$$

where: $\left(\frac{\Delta P}{L}\right)_G$ = Gas pressure drop, psi/ft (kPa/m) of bed

K = Dimensional constant = $1 / (144 \times 4.17 \times 10^8)$ (Customary)
= $1 / 1000$ (Metric)

Re_G = Gas bed Reynolds number, dimensionless
= $\frac{W_G D_p}{\mu_G (1-\varepsilon)}$

ε = Void fraction of bed, dimensionless. See **Table 1, Section VII-A** for typical values.

μ_G = Viscosity of gas at conditions, lb/ft hr (kg/ms)

ρ_G = Gas density at conditions, lb/ft³ (kg/m³)

D_p = Equivalent particle diameter, ft (m). See **Section VII-A** for definition.

W_G = Mass velocity of gas, lb/hr ft² (kg/sm²) based on reactor cross-sectional area.

Step 2: Liquid Phase Pressure Drop - Calculate $(\Delta P/L)_L$, the pressure drop which the liquid phase would develop if it were present alone, from the Ergun equation:

$$\left(\frac{\Delta P}{L}\right)_L = K \operatorname{Re}_L (150 + 1.75 \operatorname{Re}_L) \left(\frac{1-\varepsilon}{\varepsilon}\right)^3 \frac{\mu_L^2}{\rho_L D_p^3} \quad \text{Eq. (2)}$$

K = Dimensional constant = $1 / (144 \times 4.17 \times 10^8)$ (Customary)
= $1 / 1000$ (Metric)

where: $\left(\frac{\Delta P}{L}\right)_L$ = Liquid pressure drop, psi/ft (kPa/m) of bed

Re_L = Liquid bed Reynolds number, dimensionless
= $\frac{W_L D_p}{\mu_L (1-\varepsilon)}$

μ_L = Viscosity of liquid at conditions, lb/ft hr (kg/ms)

ρ_L = Liquid density at conditions, lb/ft³ (kg/m³)

W_L = Mass velocity of liquid, lb/hr ft², (kg/sm²) based on reactor cross-sectional area

Step 3: Mixed-Phase Pressure Drop - Calculate χ , the abscissa of **Figures 6 and 7**:

$$\chi = \sqrt{\left(\frac{\Delta P}{L}\right)_L} / \sqrt{\left(\frac{\Delta P}{L}\right)_G} \quad \text{Eq. (3)}$$

Then, using this value of χ , read the value of ϕ_L from **Figure 6** or **7**.

Finally, calculate the mixed-phase pressure drop, $(\Delta P/L)_{LG}$, from ϕ_L and $(\Delta P/L)_L$:

$$\left(\frac{\Delta P}{L}\right)_{LG} = \phi_L^2 \left(\frac{\Delta P}{L}\right)_L \quad \text{Eq. (4)}$$

where: $\left(\frac{\Delta P}{L}\right)_{LG}$ = Two-phase pressure drop, psi/ft (kPa/m) (correlation includes resistance to flow and static head contribution)

ϕ_L = Correlating factor for two-phase pressure drop (ordinate in **Figures 6 and 7**)

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DETAILED DESIGN PROCEDURE (Cont)

Feeds with Final Boiling Points above 800°F (700°K)

For feeds boiling above 800°F (700°K) the gas and liquid pressure drops should be calculated for each phase separately as described above. The abscissa, χ , in **Figure 8** should then be calculated from:

$$\chi = A \sqrt{\left(\frac{\Delta P}{L}\right)_L} / \left(\frac{\Delta P}{L}\right)_G \quad \text{Eq. (5)}$$

$$\text{where: } A = \begin{array}{ll} 0.21 & \text{(Customary)} \\ 1.0 & \text{(Metric)} \end{array}$$

The mixed phase pressure drop is calculated from ϕ_G , the ordinate of **Figure 8**, as:

$$\left(\frac{\Delta P}{L}\right)_{LG} = \phi_G^2 \left(\frac{\Delta P}{L}\right)_G \quad \text{Eq. (6)}$$

Heritage Mobil Parallel Flow Pressure Drop Prediction

For single-phase flow in a solids bed, whatever complex fluid mechanics may govern locally, macroscopic frictional losses follow the Ergun equation. For gas-liquid flow this equation is therefore applied to each phase individually in the void volume that phase occupies, irrespective of whether solid or fluid boundaries define the volume. This method takes the original Ergun equation as a valid description of flow in any geometry sufficiently tortuous that profile reconstruction dominates, and it applies to each phase volume unchanged. That each phase in effect reshapes the other is captured in the model by solving the gas and liquid Ergun equations simultaneously for the two phase void fractions ϵ_G and ϵ_L subject to phase pressure drop equivalence, which sum to the total flowing void. Hence the Parallel Flow model is completely defined by the following relationships:

$$\left(\frac{\Delta P}{L}\right)_G = K \left(\frac{150(1-\epsilon_G)\mu_G}{D_p} + 1.75 G_G \right) \frac{(1-\epsilon_G)}{\epsilon_G^3} \left(\frac{W_G}{\rho_G g D_p} \right) \quad \text{Eq. (7)}$$

$$\left(\frac{\Delta P}{L}\right)_L = K \left(\frac{150(1-\epsilon_L)\mu_L}{D_p} + 1.75 G_L \right) \frac{(1-\epsilon_L)}{\epsilon_L^3} \left(\frac{W_L}{\rho_L g D_p} \right) \quad \text{Eq. (8)}$$

$$\epsilon_G + \epsilon_L = \epsilon_F \quad \text{Eq. (9)}$$

$$\Delta P_G = \Delta P_L \quad \text{Eq. (10)}$$

The head term is omitted from the pressure balance since forces supporting the phases are variously translated into the packing and container walls.

It is understood that the gas and liquid void fractions in equations describe not pointwise values but rather phase partitioning in a region of packing large enough to include flowing and pressure neutral volume in proportions typical of the entire bed. Also, if the flux of either phase drops to zero the Parallel Flow treatment reduces identically to an Ergun equation for the other phase using flowing void fraction.

In practice **Eqs. 7-11** are combined into a single formula for the gas void fraction:

$$\left(\frac{150(1-\epsilon_G)\mu_G}{D_p} + 1.75 G_G \right) \frac{(1-\epsilon_G)}{\epsilon_G^3} \left(\frac{G_G}{\rho_G} \right) = \left(\frac{150[1-(\epsilon_F-\epsilon_G)]\mu_L}{D_p} + 1.75 G_L \right) \frac{[1-(\epsilon_F-\epsilon_G)]}{(\epsilon_F-\epsilon_G)^3} \left(\frac{G_L}{\rho_L} \right) \quad \text{Eq. (11)}$$

This equation converges smoothly using a Newton-Raphson method, typically in 2 to 5 iterations. A good initial value is 2/3 of the flowing void fraction. Bed pressure drop is then computed directly from equation.

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◆ S INLET DISTRIBUTORS

There are a number of inlet nozzle distributor designs that have been used throughout heritage Mobil. The design procedure for the preferred design has recently been formalized as described in **Reference 10**. **Figure 9** shows a schematic view of the base design. For evaluation of these designs or to develop a new design see **Reference 10** and contact the Chemical Engineering Technology Section of EMRE for the spreadsheet used to calculate required dimensions and pressure drop. This inlet nozzle distributor reduces the kinetic energy of the incoming liquid which results in a more uniform liquid level on the liquid distributor tray below. This has particular benefits in retrofit designs in which there is insufficient space available for a calming tray.

An alternative design has been used throughout heritage Exxon units. The vapor-liquid inlet pipe enters the vessel through the top head and extends into the reactor at least 3 in. (76 mm) below the top head (or lining in case of a lined vessel). An impingement baffle with a diameter twice that of the inlet pipe is attached by baffle supports to the end of the inlet pipe. The baffle is required to break up high velocity currents from the inlet piping. The distance between the baffle and the bottom of the inlet pipe is such that this open area is at least equal to the cross-sectional area of the inlet pipe. The vertical distance between the baffle and any lining in the top head should be at least 9 in. (229 mm). The baffle plate should be perpendicular to the inlet pipe to within 1/900 of the pipe diameter. One ft (305 mm) should be provided between the inlet nozzle baffle and the highest projection on the top distributor tray.

Liquid and gas are distributed uniformly across the bed by a distributor tray. If the distributor tray is not removed for catalyst loading and leveling, an accessway must be provided in the tray. In this case, the distance between the bottom of the tray and top of the bed (including inerts) should be 3 ft (914 mm) to allow room for leveling the catalyst bed. If the tray is removed, then a minimum distance of 1.5 ft (457 mm) must be provided from the bottom of the tray support beams to the top of the inerts.

Spray Nozzle Distributor Tray (Figure 1) - Design of a mixed-phase spray nozzle distributor requires use of a standard, commercially available, spiral type spray nozzle, as shown in **Figure 1**, that is normally used for single liquid phase spray distribution. Use of these nozzles for mixed-phase service is an ExxonMobil Proprietary Design. Design of a spray nozzle distributor requires use of an EXCEL spreadsheet that contains a neural network model to predict pressure drop. Contact the Chemical Engineering Technology Section of EMRE to obtain this spreadsheet. The program inputs required are gas and liquid volumetric flow rates and densities, reactor inside diameter, and locations and depth of beams. The spreadsheet also contains a number of macros that will greatly facilitate the design process. These macros calculate the number of nozzles and locations for a given pitch, allow adjustments of the nozzle positions, and write the output file required for input to a computer model (see below) used to calculate the liquid distribution on the catalyst bed. The logic of the design process is iterative as follows:

1. Determine approximate number of nozzles required.
2. Determine nozzle size based on pressure drop requirements.
3. Determine nozzle locations for standard pitch layout.
4. Adjust nozzle locations for clearance and eliminate those too close to thermowells or the wall.
5. Return to first step and redo steps as needed.

The following guidelines should be followed for the design.

Determination of the Number of Nozzles: The number of spray nozzles should be between 1.5 and 3 nozzles/ft² (16 and 32 nozzles/m²) of reactor cross-sectional area to obtain a satisfactory liquid distribution. Most applications will fall between 2.0 and 2.5 nozzles/ft² (22 and 27 nozzles/m²). For low gas flow cases, $pv^2 < 500$ lb/ft² (744 kg/ms²) through each nozzle and liquid mass velocities below 1500 lb/hr ft² (2.04 kg/sm²), there should be at least 2 nozzles/ft² (22 nozzles/m²) and preferably 3 nozzles/ft² (32 nozzles/m²). The velocity v in ft/s (m/s) used in the pv^2 parameter is the superficial gas velocity through the nozzles. It is calculated by dividing the total gas volumetric flow rate by the total cross-sectional throat area of all the nozzles. For high liquid mass velocities above 3000 lb/hr ft² (4.07 kg/sm²), and for some fouling and all quench box operations, 1.5 nozzles/ft² (16 nozzles/m²) will be adequate. Larger vessels will tend to have fewer nozzles/area. Smaller vessels will have more nozzles/area because the cost of the nozzles is a lower fraction of the total cost. The final choice should be made based on the ability to obtain a uniform liquid distribution on the top of the catalyst bed.

A number of factors enter into the determination of the exact number of nozzles. The most important of these is the layout of the support beams. For a given beam layout a certain pitch will usually fit the design cleanly between beams, reducing the number of adjustments to be made for clearance and reducing the liquid spray off the beams and improving liquid distribution on the top of the catalyst bed. For grass roots designs the mechanical supports can be designed to accommodate an optimal nozzle layout.

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Determination of Nozzle Size: Once the number of nozzles has been set, the nozzle size is selected to meet pressure drop requirements. The list of available nozzles (including sizes, spray angle, and manufacturer designations) is specified in the spreadsheet design program. A minimum of 10 in. (254 mm) of water pressure drop is desirable to obtain good liquid dispersion. If this cannot be achieved while meeting the nozzle density requirement, then a minimum gas pv^2 of 500 lb/ft s² (744 kg/m/s²) (superficial gas velocity based on nozzle throat area) is required to obtain good liquid dispersion. For new designs a calculated tray pressure drop (nozzle plus liquid head) of 1.0 psi (6.9 kPa) and a tray design for 2.0 psi (13.8 kPa) should be used. Mechanical constraints usually limit the maximum pressure drop available across trays to about 1 - 2 psi (6.9 - 13.8 kPa). The nozzle pressure drop, as predicted by the neural net model in the spreadsheet must be added to the liquid head on the spray tray and the total should be limited to about 1/2 the maximum allowable pressure drop. For retrofit designs a mechanical evaluation of the existing beams and support ring is required to determine the maximum loading allowable. For both new and retrofit designs the emergency depressurization conditions must be reviewed to determine their effects on tray loading.

Determination of Nozzle Locations: A triangular pitch layout should be used with the pitch chosen to give the correct number of nozzles. Nozzles should be no closer than 3 in. (75 mm) from the edges of beams, and 6 in. (152 mm) from the wall and thermowell. Nozzle locations should be adjusted to maintain a good spray pattern. Occasionally, some nozzles may need to be removed from the layout, e.g., when a nozzle is directly located on a thermowell. Following these guidelines, wide angle spray nozzles (120°) should be used for catalyst beds 12 - 18 in. (305 - 457 mm) below the tray (retrofit design). Moderate angle nozzles (90°) should be used for beds 18 - 27 in. (305 - 686 mm) below the tray. Narrow angle (60°) nozzles should be used for beds 27 - 36 in. (686 - 914 mm) below the tray. These guidelines refer to start of run loading specifications for dense loaded catalyst. For loosely loaded catalyst where some bed slumping is expected, the next narrower angle spray nozzle may be required. The spray patterns from the individual nozzles overlap to produce a uniform spray pattern, therefore the positions of the nozzles to the nearest 1/2 in. (12.7 mm) should be specified in (x, y) coordinates and should be reported to the contractor.

A computer model has been developed to estimate the liquid distribution on the catalyst bed from the spray nozzles. Contact the Chemical Engineering Technology Section of EMRE for use of this model to ensure that a good liquid distribution will be obtained.

Chimney Design: The design of the slotted chimneys for use in the spray trays is similar to the design for the chimneys in the short tube-slotted chimney trays (see below). The slots should start at least 2 in. (51 mm) above the tray deck. For fouling operations where foulant accumulates on the tray this height should be increased to 4 in. (102 mm) or higher if experience indicates larger amounts of fouling. Slot width should be greater than 1/4 in. (6.4 mm). Slot width is determined using **Eq. (14)**. For fouling operations use a wider slot width, preferably 1/2 in. (12.7 mm). For slot widths greater than 1/2 in. (12.7 mm) use more than one slot per chimney as required. Total slot height should be approximately 2 in. (51 mm) higher than that required for the maximum expected flow rate.

- ▶ **Chimney Downcomer Spray Tray (Figure 2)** - Current practice is to standardize designs using downcomer tubes 1 1/2 in. or 2 in. NPS sch 40 pipe and to use two overflow slots of 1/2 in. width and 1 in. height. This leaves hole size(s), height(s), and number(s) as well as the number and height of downcomers to complete the geometry definition. The spray angle increases and liquid droplet size decreases as the vapor velocity through the downcomer increases, i.e., with decreased number of downcomers. Therefore a proper balance between distributor density, spray pattern, and liquid atomization must be achieved to most effectively utilize the catalyst. For new designs a layout pitch can be estimated assuming a spray angle of 30 degrees. A minimum of four downcomer tubes per sq. ft. is recommended. Detailed design procedures are available in a spreadsheet and in an accompanying manual by contacting the Chemical Engineering Technology Section of EMRE. Results of the cold flow studies used to develop these parameters can be found in **Reference 9**. Familiarization with these results is highly recommended for effective use of the procedures.

Downcomers should not be capped if a rough distributor, or other means for preventing liquid raining on the downcomers, is provided upstream of the downcomer tray. If capping is specified a minimum clearance of 2 in. or one downcomer ID, whichever is larger, should be allowed between the cap and the downcomer top rim.

Short Tube-Slotted Chimney Tray (Figure 3) - The following guidelines should be met in designing the short tube slotted chimney distributor. Generally, a design with about 50% of the flow through the tubes and 50% of the flow through the slotted chimneys is desirable, since this will give the most uniform distribution with only moderate sensitivity to variations in level. Designs with somewhat more flow through the tubes (up to 60%) can be used, but designs with more than 50% of the flow through the slotted chimneys should not be specified, because they will be too sensitive to small differences in level.

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In specific situations it may be desirable to deviate from the dimensions given above, but the general design parameters should be maintained. In particular for designs with blocked operations or large turndown requirements, deviations from the above design procedure may be required. In general, these trays should be designed such that the liquid level at 40% of the lowest throughput is 1-1/2 in. (38 mm) above the short tubes, and at the highest rate the liquid level is 1 in. (25 mm) above the slots. In these designs h_1 will be greater than 1-1/2 in. (38 mm), and more than 50% of the liquid will flow through the tubes at maximum throughput. It may also be necessary to use tubes longer than 2-1/2 in. (63.5 mm) in these situations. In some blocked operations, sets of two different tube heights have been used to help assure even distribution over the full liquid flow range. The tube height difference should be at least 1-1/2 in. (38 mm). The consequence of using an intermediate tube length is to stack the height of the slotted chimney 1-1/2 in. (38 mm) higher than normal.

The number and diameter of the slotted chimneys should be checked to ensure that gas pressure drop across the tray at maximum vapor rate (treat gas plus vaporized feed) is less than 5% of the liquid head on the tray. If this condition is not satisfied, the number or diameter of the chimneys should be increased. Pressure drop across the tray can be calculated from:

$$(\Delta P)_{\text{tray}} = K \rho_g v_c^2 \quad \text{Eq. (15)}$$

where:

$(\Delta P)_{\text{tray}}$	=	Tray pressure drop, psi (kPa)
v_c	=	Vapor velocity through chimney, ft/s (m/s)
ρ_g	=	Vapor density lb/ft ³ (kg/m ³)
K	=	0.00031 (Customary)
	=	0.00144 (Metric)

Normally, if the tray pressure drop constraint is satisfied at maximum vapor rate, it will be satisfied at any lower throughput. However, this may not hold true in some blocked operations or in situations where the liquid feed rate is lowered without reducing treat gas rate. In these cases the tray pressure drop constraint should be checked at all planned operating conditions.

Bubble Cap Distributor (Figure 4) - The bubble cap distributor has been used in mixed-phase packed bed reactors, primarily for its heat transfer capability. The design utilizes the gas pressure drop across the distributor to force the liquid uniformly through the caps.

Bubble caps are placed on a 6 in. (152 mm) to 10 in. (254 mm) triangular pitch with the centerline of the outside cap 4-1/2 in. (114 mm) from the vessel wall. Cap dimensions are given in **Figure 4**. Because of their inability to distribute liquid uniformly at low pressure drop and our lack of experience with these at high pressure drops, the bubble cap distributor should be used only when:

$$K_1 \geq V_{BC} \sqrt{\rho_G} \geq K_2 \quad \text{Eq. (16)}$$

where:

V_{BC}	=	Actual gas flow per cap, ft ³ /hr (m ³ /s)
K_1	=	1000 (Customary)
	=	0.0315 (Metric)
K_2	=	300 (Customary)
	=	0.0094 (Metric)

Although bubble cap trays are no longer recommended a more detailed study of their operation is now available in **Reference 9**. For analysis of existing operations a design tool for bubble cap rating is available in the spreadsheet design tool for chimney downcomer spray trays. Contact the Chemical Engineering Technology Section of EMRE for this tool.

LIQUID REDISTRIBUTION

Liquid redistributors, when used, should be the same type as the inlet distributor. The bottom of the upper bed section and the redistributor plate should be separated by sufficient distance to permit installation of inter-bed hardware and the operation of a grid trap (see below). A 3 in. to 6 in. (76 mm to 152 mm) layer of inert balls should be placed on top of the catalyst in the lower section and at the bottom of the upper bed section. The inert ball diameter should be a maximum of 4 times the catalyst particle size. Multiple layers of inert balls may be required. See **Section VII-B** for criteria on minimum inert ball size on top of catalyst in the lower bed.

The bed support grid should have more than 50% open area and may include a grid trap designed to open downward. When used, this grid trap should contain a release mechanism, which is operable from outside the vessel through an open manhead.

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QUENCH HARDWARE

The swirl cap design or the spider vortex design should be used in all new applications. The previous design, although no longer recommended, is also described below.

SWIRL CAP QUENCH ASSEMBLY DESIGN (REFERENCE 6)

The design of the improved quench box has been facilitated by several Lotus PC spreadsheets. The use of these spreadsheets is referred to in the sections which follow and is described in detail in **Tables 1** through **5**. The spreadsheets are referred to collectively as Program CPEE090. Installation and startup instructions for this program are given in **Table 6**. The program may be ordered from EMRE Manuals Distribution using the form shown in **Table 7**.

Figure 11 shows an isometric view of the new quench assembly. **Figure 12** shows a vertical cut-away of the assembly. These figures show a bubble cap tray for liquid redistribution. A spray tray could also be used and is preferred in a new installation.

The quench flow is directed through a multi-holed sparger transverse to the flow of hot reactants from the bed above. The cold quench jets entrain and mix with part of the hot reactant process flow from the bed above. This comprises the first quench mixing stage. Two alternative sparger designs may be used for quench injection. These are shown in **Figures 13** and **14**.

The partially pre-mixed quench and process flows are directed radially inward toward the mixing cap located at the center of the top tray. The angled vanes in the swirl mixing cap give a centrifugal swirling motion to the gas and liquid within the cap. This mixture flows radially inward then downward toward the mixing table. When the volumetric liquid/gas ratio is greater than 0.1 the slotted swirl cap shown in **Figure 15** is recommended. Otherwise the vane swirl cap shown in **Figure 16** should be used.

As the flow moves outwardly over the mixing table, it rolls off the edge and attains a torodial motion, further enhancing mixing. The mixed flow then moves outward toward the peripheral outlet holes on the middle tray. The flow on the middle tray is evenly split by the peripheral outlet holes and is projected downward as spraying jets toward the annular donut baffle. The flow is then redirected toward the center but somewhat off the radial lines to create another swirl pattern below the middle tray by the flow redirection baffles. These baffles form horizontal jets issuing inwardly along the annular donut baffle. Two alternative designs for the flow redirection baffles are shown in **Figures 17** and **18** although the design in **Figure 17** is preferred. The combination of the flow redirection baffles and the annular donut baffle helps to form the jets and also to place the downward liquid flow at the area center of the bubble cap liquid redistribution tray below the quench assembly. This minimizes flow disturbances and hydraulic gradients on the tray below.

A bubble cap or spray liquid redistribution tray serves to further mix the vapor and liquid flows in addition to distributing the mixed phase flow evenly across the top of the next bed. The design of the liquid redistribution tray has been previously described in this section.

S DESIGN OF THE QUENCH INJECTION SPARGER FOR SWIRL CAP QUENCH ASSEMBLY

Single-Phase Quench Fluid

When the quenching medium is a single phase (all vapor or all liquid) the preferred sparger design is the Circular Pan Sparger (**Figure 13**). The design procedure for this sparger is described in **Table 1**.

In some retrofit applications, it may not be possible to use the circular pan sparger because of space limitations or turnaround considerations. For these situations an alternative multi-holed, bayonet sparger may be used. Although somewhat less symmetric than the circular pan sparger, the bayonet sparger apportions flow uniformly using a varying hole size along the sparger axis. **Table 2** is a user's guide for a Lotus Spreadsheet Program PIPESPG which calculates the hole sizes for the bayonet sparger. The hole size estimates are valid even for situations in which the allowable pressure drop across the holes is low relative to the fluid kinetic energy in the pipe.

It is important that the sparger holes for either design are not obstructed by the catalyst bed or pipe supports. In the case of the bayonet sparger, expansion of the pipe during heat up should be considered when designing pipe supports.

Two-Phase Quench

Quenching with a mixed-phase stream can reduce gas compression investment and operating cost. It can also debottleneck units that are limited by compression capacity. While spargers have been designed for two phase flows there is insufficient data available to develop generalized design tools. For two phase sparger designs consult with the Chemical Engineering Technology Section of EMRE.

Design of the Swirl Cap

The swirl mixing cap on the top tray provides the majority of the mixing. Most of the pressure drop over the quench assembly is expended through the swirl mixing cap. A pressure drop of 3 psi (20.7 kPa) across the top tray is a reasonable design basis. However, lower pressure drops are often feasible. It is important that both the process and mechanical support designs are adequate for the chosen pressure drop. For existing units, both the pressure drop and tray separation distance are set by the existing tray supports.

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High Liquid / Vapor Ratio

If the volumetric liquid/vapor ratio of the main process flow exceeds 0.1 the slotted swirl cap (**Figure 15**) should be used. The internal cylindrical baffle in this design promotes increased liquid mixing and gas/liquid contacting. The design procedure for this device is incorporated in a Lotus Spreadsheet QBSSC. The procedure for using this program is shown in **Table 3**.

Low Liquid / Vapor Ratio

If the volumetric liquid/vapor ratio of the main process flow does not exceed 0.1 the vane swirl cap (**Figure 16**) should be used. The design procedure for this device is incorporated in a Lotus Spreadsheet QBVSC. The procedure for using this program is shown in **Table 4**.

Design Procedure for the Middle Tray

The middle tray design consists of three parts; determination of the distance between the top tray and the middle tray, sizing the cylindrical mixing table, and determining the size and number of outlet holes. The middle tray layout is shown in **Figure 12**.

Separation Distance between Top Tray and Middle Tray

Mixing performance is a strong function of the ratio of the tray separation distance (**Dimension A in Figure 12**) to the swirl cap hole diameter (**Dimension D in Figures 15 and 16**). For grass roots units this ratio should be no less than 0.33 but preferably 0.5. In retrofit applications where the tray separation distance is fixed, the hole size should be reduced to the minimum value consistent with the allowable pressure drop. Consult with the Chemical Engineering Technology Section of EMRE if this results in a ratio that is less than 0.33 or if mixed phase quench is used.

Mixing Table Sizing

The mixing table is situated on the middle tray directly below the hole in the swirl mixing cap (**Figure 12**). The diameter of the mixing table is exactly equal to the swirl mixing cap hole diameter. The height of the table is preferably 1/2 the tray separation distance but may be reduced to 1/3 the tray separation distance if necessary to minimize reactor straight side.

Number and Size of Outlet Holes

The outlet hole centers are equally spaced on a circle at a distance from the reactor wall (**Dimension A in Figures 17 and 18**) somewhat larger than the tray support ring. **Table 5** describes the procedure for calculating the number of holes and hole size using the Lotus Spreadsheet MIDTRAY. The objective is to provide a minimum of 0.25 psi (1.7 kPa) pressure drop for uniform flow distribution among the holes.

Design of the Annular Donut Baffle

The donut baffle prevents the jets issuing from the outlet holes from disrupting the flow distribution on the liquid redistribution tray. It also redirects the flow leaving the outlet holes in a radial swirling pattern to further mix the quench and process flow. Two designs are available for this baffle as shown in **Figures 17 and 18**. The design of the annular donut baffle consists of three parts: the flow redirection baffle, the separation distance between the annular donut baffle and the middle tray, and the width of the horizontal section of the annular donut baffle.

Flow Redirection Baffle

The preferred method of redirecting the flow below the middle tray is the "u-shaped" baffle shown in **Figure 17**. This baffle is situated directly below each outlet hole. The radius of curvature of the bend is exactly that of the outlet hole so that the baffle does not interfere with the vertical downward flow path leaving the hole. The centerline of the "u-shaped" baffle is directed at an angle to the radial line connecting the center of the reactor and the center of the outlet hole. This swirl angle is calculated such that the jets issuing from each flow redirection baffle do not impinge on the neighboring "u-shaped" baffles.

In reactors with solid, chordal tray support members (as opposed to chordal trusses or radial supports extending from the tray support ring) an alternative flow redirection baffle, the notched tube, is recommended. The inside diameter of this tube is exactly equal to the diameter of the outlet hole. The tube extends vertically downward below the middle tray support members and touches the horizontal baffle below. The bottom end of the tube is notched as shown in **Figure 18** to allow the flow to exit. The notch extends horizontally to the diameter of the tube.

The Spreadsheet MIDTRAY calculates the number of outlet holes, and the angle of orientation to the radial line (**Dimension F**) in **Figure 18**.

Dimensions of the Annular Donut Baffle

The horizontal baffle is sized such that the inner edge is located at a diameter that is 0.707 times the reactor inside diameter (**Dimension B in Figures 17 and 18**). The baffle completely covers the bottom of the flow redirection baffle such that the entire flow is directed horizontally and toward the inner edge of the baffle. The vertical outer ring is situated between the reactor wall and the redirection apparatus at the outer edge of the horizontal baffle such that the clearance between the ring and the apparatus is the minimum required for ease of fabrication. The outer ring extends upward at least half of the separation distance between the horizontal baffle and the splash tray.

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Separation Distance Between the Horizontal Baffle and the Splash Tray

For designs using the "u-shaped" flow redirection baffle, the separation distance is equal to the vertical height of the baffle (**Dimension C** in **Figure 17** or **Dimension G** in **Figure 18**). (See **Table 5** for **Dimension C**) For designs using the notched tube the separation distance is the distance the solid tray support member extends below the splash tray plus the vertical notch height.

Flow Obstruction Criteria

Due to the importance of centrifugal and swirling motion to the mixing in the quench assembly, it is important to minimize flow obstructions. Supporting members and trusses for all decks and the annular baffle should have at least 25% open area for flow.

Leakage Criteria for Quench Assembly/Tray Levelness

All leakage constitutes bypassing. Therefore, leakage should be prevented since it can cause localized hot spots in the catalyst bed below. Tray panel sections should be gasket sealed or seal welded with each other and with the support ring on the vessel wall. Total tray bypassing due to leakage should not exceed 1% of the process flow. Leakage should be tested in situ with a water test. Refer to **GP 5-2-2** for details regarding leak testing. **GP 5-2-2** also contains specifications for required tray levelness.

Leak Test Criteria

The specific conditions for the test depend on the reactor conditions, fluid properties, and tray pressure drops. This is discussed in more detail below.

Two-Phase Flow (Liquid/Vapor)

For two phase systems (vapor and liquid present under all conditions) the leakage through the trays is mainly liquid. Therefore, the maximum allowable tray leakage rate is 1% of the liquid flow to the tray at process conditions. This can be translated into an allowable water leakage rate using the following equation.

$$L_{\text{water}} = K \frac{\Delta P_{\text{water}}}{\Delta P_{\text{tr}}} \mu_{\ell} L_{\text{tray}} \quad \text{Eq. (17)}$$

where:	L_{water}	=	Maximum allowable water leak test rate, gpm/ft ² (m ³ /s/m ²)
	L_{tray}	=	1% of design rate @ conditions, gpm/ft ² (m ³ /s/m ²)
	ΔP_{water}	=	Depth of water on tray during leak test, in. (m)
	ΔP_{tr}	=	Pressure drop across tray @ conditions, in. of H ₂ O, (m of H ₂ O)
	μ_{ℓ}	=	Viscosity of liquid @ conditions, cp (kg/m/s)
	K	=	1 (Customary) 1000 (Metric)

Vapor Phase Flow

For vapor phase systems the leakage through the trays is vapor. Therefore, the maximum allowable tray leakage rate is 1% of the total flow to the tray at process conditions. This can be translated into an allowable water leakage rate using the following equation:

$$L_{\text{water}} = K \frac{\Delta P_{\text{water}}}{\Delta P_{\text{tr}}} \mu_{\text{v}} V_{\text{tray}} \quad \text{Eq. (18)}$$

where:	L_{water}	=	Maximum allowable water leak test rate, gpm/ft ² (m ³ /s/m ²)
	V_{tray}	=	1% of design rate @ conditions, ft ³ /min/ft ² (m ³ /s/m ²)
	ΔP_{water}	=	Depth of water on tray during leak test, in. (m)
	ΔP_{tr}	=	Pressure drop across tray @ conditions, in. of H ₂ O (m of H ₂ O)
	μ_{v}	=	Viscosity of vapor @ conditions, cP (kg/m/s)
	K	=	7.48 (Customary) 1000 (Metric)

Thermowell Guides

Vertical thermowells protruding through the quench assembly should be equipped with thermowell guides. These guides should be essentially the height of the box for the middle tray and the height of the mixing cap for the top deck. The gap between the thermowell guide and the thermowell should be the minimum necessary to allow for expansion.