CL5005 REACTION ENGINEERING

MODULE

II

DESIGN OF IDEAL REACTORS

Lecture Notes:
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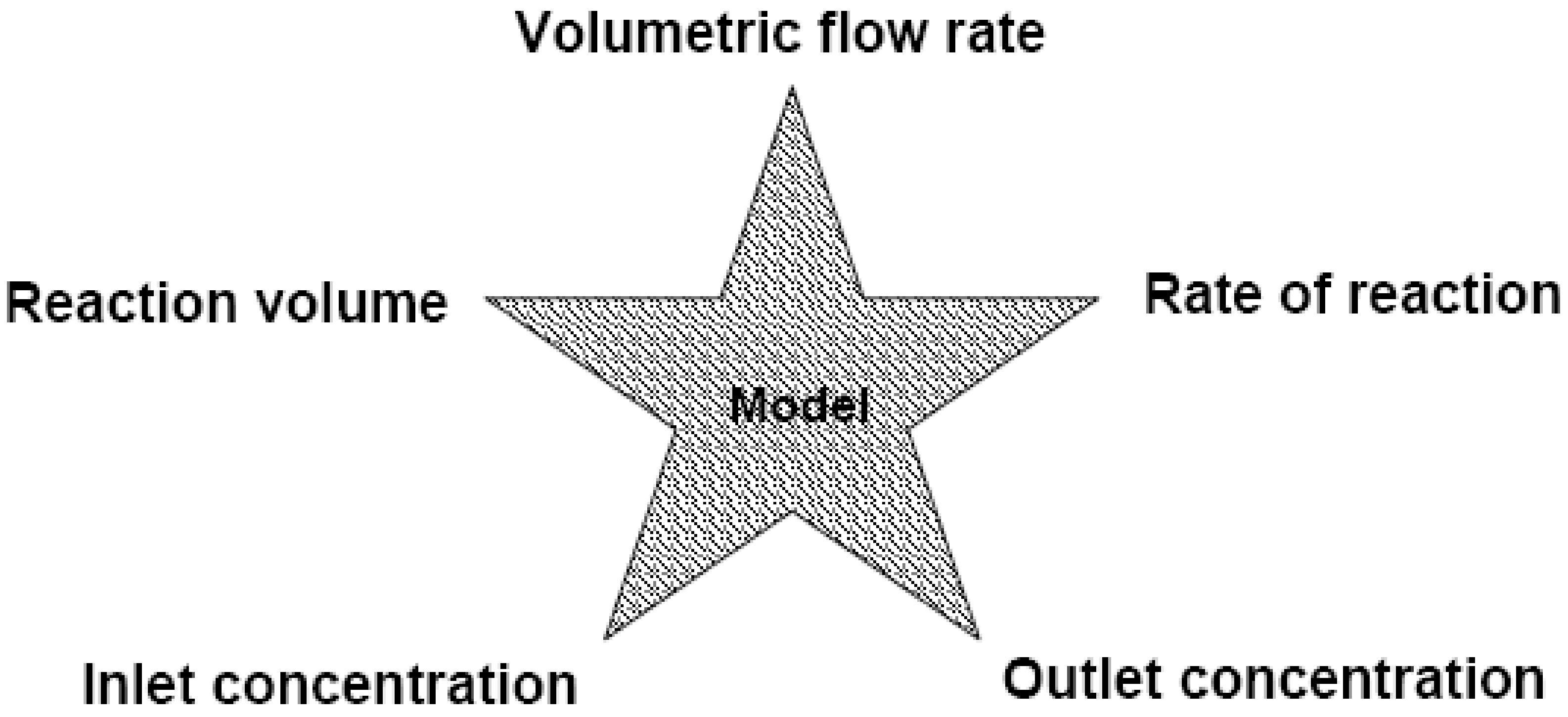
Introduction To Reactor Design

- In reactor design we want to know what size and type of reactor and method of operation are best for a given job. Because this may require that the conditions in the reactor vary with position as well as time, this question can only be answered by a proper integration of the rate equation for the operation. This may pose difficulties because the temperature and composition of the reacting fluid may vary from point to point within the reactor, depending on the endothermic or exothermic character of the reaction, the rate of heat addition or removal from the system, and the flow pattern of fluid through the vessel. In effect, then, many factors must be accounted for in predicting the performance of a reactor. How best to treat these factors is the main problem of reactor design.
- In this module, several ideal types of reactors are discussed based on two modes of operation (batch and continuous), and ideal flow patterns (back-mix and tubular) for the continuous mode.
- four types of ideal reactors are : (1) Batch reactor (BR), based on complete mixing; (2) Continuous-flow stirred tank reactor (CSTR), based on back-mix flow; (3) Plug-flow reactor (PFR), based on plug flow; and (4) Laminar-flow reactor (LFR), based on laminar flow.

Introduction To Reactor Design

Reactor Design



Volumetric flow rate

Reaction volume

Rate of reaction

Model

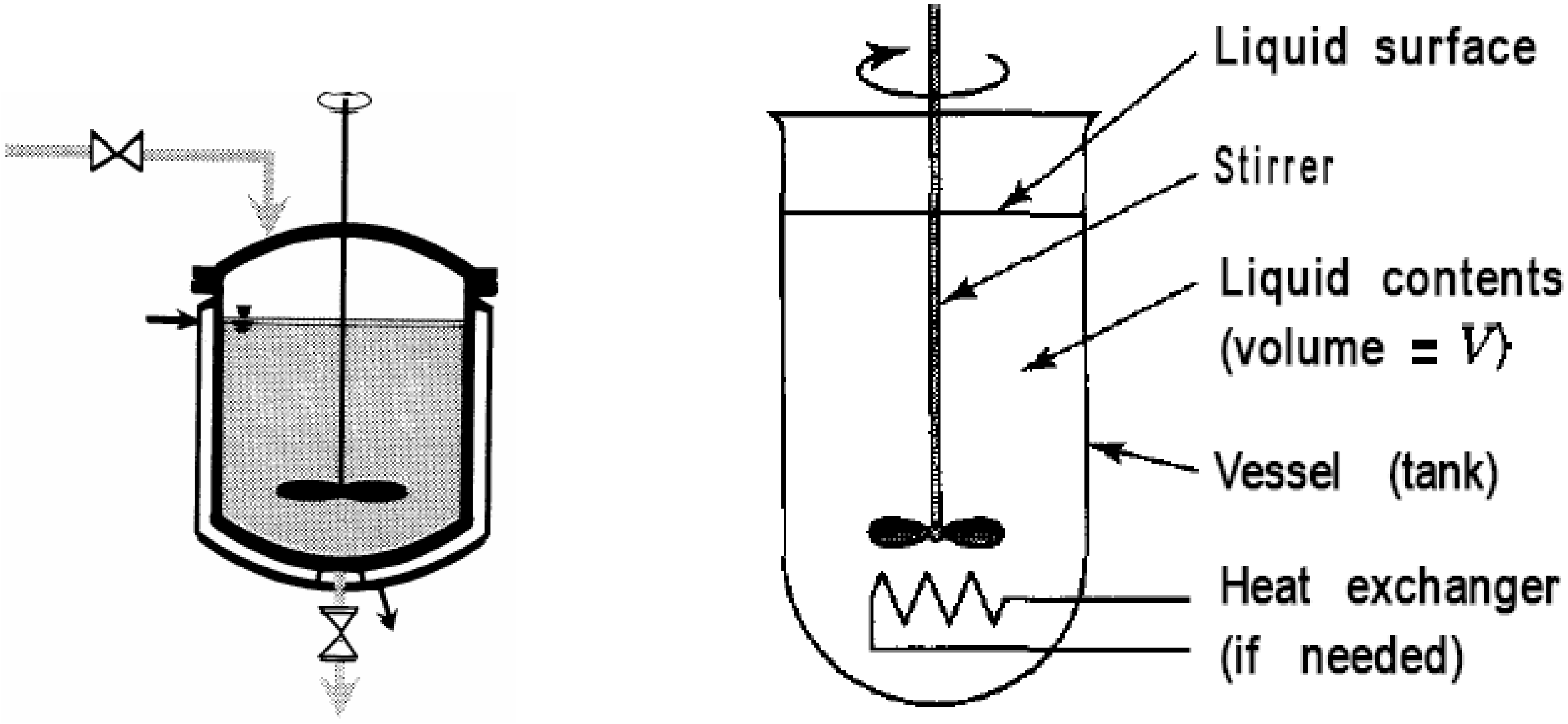
Inlet concentration

Outlet concentration

Outlet = f(inlet, kinetics, contacting pattern)

Batch Reactor (BR)

- A batch reactor (BR) is sometimes used for investigation of the kinetics of a chemical reaction in the laboratory, and also for larger-scale (commercial) operations in which a number of different products are made by different reactions on an intermittent basis.



Batch reactor (schematic, liquid-phase reaction)

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- Characteristics of BR

1. Each batch is a closed system.
2. The total mass of each batch is fixed.
3. The volume or density of each batch may vary (as reaction proceeds).
4. The energy of each batch may vary (as reaction proceeds); for example, a heat exchanger may be provided to control temperature, as indicated in above Figure .
5. The reaction (residence) time t for all elements of fluid is the same.
6. The operation of the reactor is inherently unsteady-state; for example, batch composition changes with respect to time.
7. Point (6) notwithstanding, it is assumed that, at any time, the batch is uniform (e.g., in composition, temperature, etc.), because of efficient stirring.

Performance equation of BR

- In a batch reactor, since the composition is uniform throughout at any instant of time, we may make the accounting about the whole reactor. Noting that no fluid enters or leaves the reaction mixture during reaction,
- Material balance with respect to reactant A(say) is as follows:

$$\overset{=0}{\cancel{\text{input}}} = \overset{=0}{\cancel{\text{output}}} + \text{disappearance} + \text{accumulation}$$

or

$$+\begin{pmatrix}\text{rate of loss of reactant A}\\ \text{within reactor due to}\\ \text{chemical reaction}\end{pmatrix} = -\begin{pmatrix}\text{rate of accumulation}\\ \text{of reactant A}\\ \text{within the reactor}\end{pmatrix}$$

$$\begin{matrix}\text{disappearance of A}\\ \text{by reaction,}\\ \text{moles/time}\end{matrix} = (-r_A)V = \left(\frac{\text{moles A reacting}}{(\text{time})(\text{volume of fluid})}\right)(\text{volume of fluid})$$

$$\begin{matrix}\text{accumulation of A,}\\ \text{moles/time}\end{matrix} = \frac{dN_A}{dt} = \frac{d[N_{A0}(1-X_A)]}{dt} = -N_{A0}\frac{dX_A}{dt}$$

Or

$$(-r_A)V = N_{A0}\frac{dX_A}{dt} \quad \text{there fore} \quad t = N_{A0}\int_0^{X_A}\frac{dX_A}{(-r_A)V}$$

Continued…

- The performance equation of BR with constant density system is

$$t = N_{A0} \int_0^{X_A} \frac{dX_A}{(-r_A)V}$$

- For all reactions in which the volume of reacting mixture changes proportionately with conversion, such as in single gas-phase reactions with significant density changes, above equation becomes

$$t = N_{A0} \int_0^{X_A} \frac{dX_A}{(-r_A)V_0(1 + \varepsilon_A X_A)} = C_{A0} \int_0^{X_A} \frac{dX_A}{(-r_A)(1 + \varepsilon_A X_A)}$$

Space-Time and Space-Velocity

- SPACE-TIME

$$\tau = \frac{1}{s} = \left(\begin{array}{c}\text{time required to process one}\\ \text{reactor volume of feed measured}\\ \text{at specified conditions}\end{array}\right) = [\text{time}]$$

- SPACE-VELOCITY

$$s = \frac{1}{\tau} = \left(\begin{array}{c}\text{number of reactor volumes of}\\ \text{feed at specified conditions which}\\ \text{can be treated in unit time}\end{array}\right) = [\text{time}^{-1}]$$

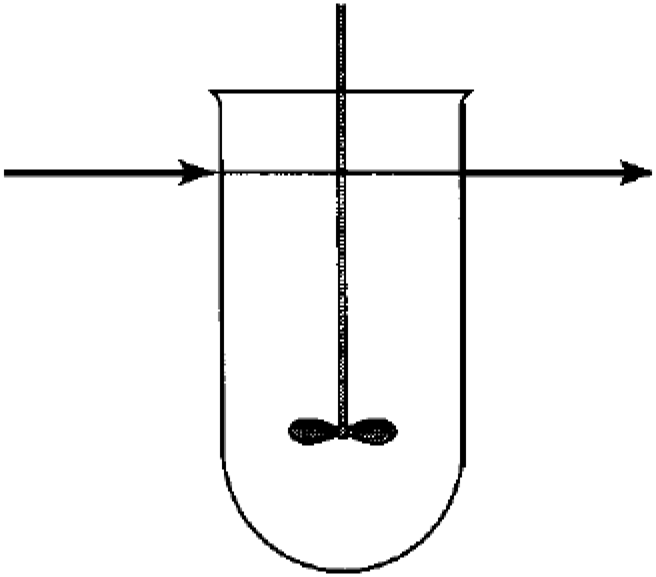
- Again

$$\tau = \frac{1}{s} = \frac{C_{A0}V}{F_{A0}} = \frac{\left(\frac{\text{moles A entering}}{\text{volume of feed}}\right)(\text{volume of reactor})}{\left(\frac{\text{moles A entering}}{\text{time}}\right)}$$

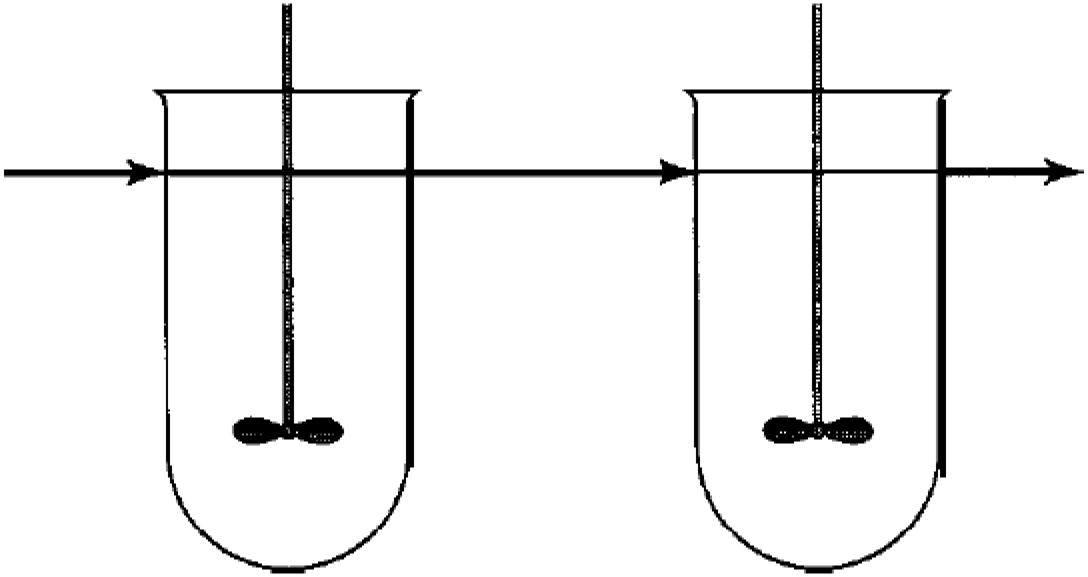
$$= \frac{V}{v_0} = \frac{(\text{reactor volume})}{(\text{volumetric feed rate})}$$

Continuous-flow stirred tank reactor (CSTR)

- A continuous stirred-tank reactor (CSTR) is normally used for liquid-phase reactions, both in a laboratory and on a large scale. It may also be used, however, for the laboratory investigation of gas-phase reactions, particularly when solid catalysts are involved, in which case the operation is batch-wise for the catalyst.
- Stirred tanks may also be used in a series arrangement (e.g., for the continuous copolymerization of styrene and butadiene to make synthetic rubber).



Single CSTR



2 CSTR in series

Continued…

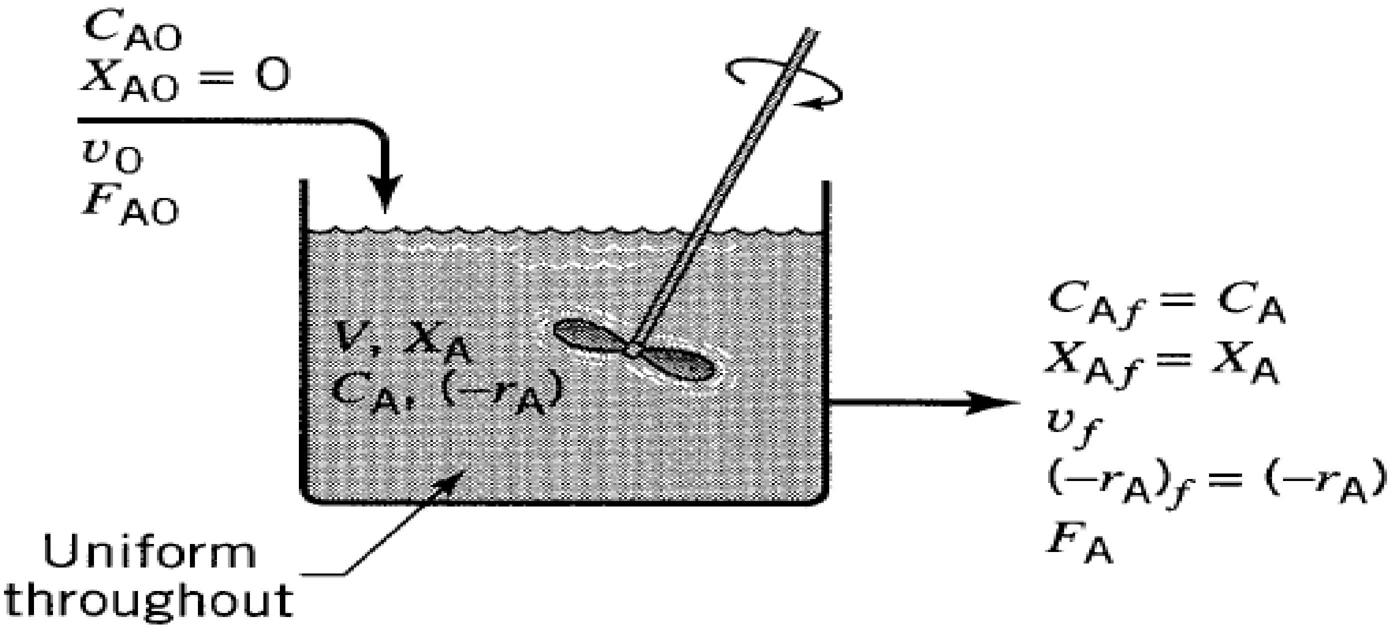
- Characteristics of CSTR

1. The flow through the vessel(s), both input and output streams, is continuous but not necessarily at a constant rate.
2. The system mass inside each vessel is not necessarily fixed.
3. The fluid inside each vessel is perfectly mixed (back-mix flow, BMF), and hence its properties are uniform at any time, because of efficient stirring.
4. The density of the flowing system is not necessarily constant; that is, the density of the output stream may differ from that of the input stream.
5. The system may operate at steady-state or at unsteady-state.
6. A heat exchanger may be provided in each vessel to control temperature

Performance equation of CSTR

- The fluid inside the vessel is uniformly mixed (and hence elements of fluid are uniformly distributed), all fluid elements have equal probability of leaving the vessel in the output stream at any time. Therefore, the output stream has the same properties as the fluid inside the vessel. If there is a step-change across the inlet in any property of the system that changes from inlet to c
- Material

$$\text{input} = \text{output} + \text{disappearance by reaction} + \overset{=0}{\cancel{\text{accumulation}}}$$



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- **As referred to the above figure**

$$\text{input of A, moles/time} = F_{A0}(1 - X_{A0}) = F_{A0}$$

$$\text{output of A, moles/time} = F_A = F_{A0}(1 - X_A)$$

$$\begin{matrix}\text{disappearance of A} \\ \text{by reaction,} \\ \text{moles/time}\end{matrix} = (-r_A)V = \left(\frac{\text{moles A reacting}}{\text{(time)(volume of fluid)}}\right)\left(\begin{matrix}\text{volume of} \\ \text{reactor}\end{matrix}\right)$$

Or

$$F_{A0}X_A = (-r_A)V$$

Therefore

$$\boxed{\begin{aligned} \frac{V}{F_{A0}} &= \frac{\tau}{C_{A0}} = \frac{\Delta X_A}{-r_A} = \frac{X_A}{-r_A} \\ \tau = \frac{1}{s} &= \frac{V}{v_0} = \frac{VC_{A0}}{F_{A0}} = \frac{C_{A0}X_A}{-r_A} \end{aligned}}$$

$$\text{any } \varepsilon_A$$

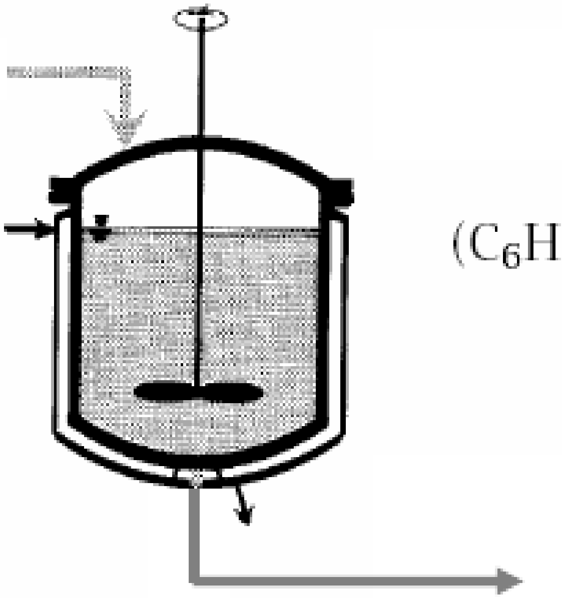
Continued….

Order of reaction	ε_A	Performance equation
1st order	0	$C_A/C_{A0} = 1 - X_A$ $k\tau = \dfrac{X_A}{1 - X_A} = \dfrac{C_{A0} - C_A}{C_A}$
1st order	Any value	$\dfrac{C_A}{C_{A0}} = \dfrac{1 - X_A}{1 + \varepsilon_A X_A}$ $k\tau = \dfrac{X_A(1 + \varepsilon_A X_A)}{1 - X_A}$
2nd order	0	$C_A = \dfrac{-1 + \sqrt{1 + 4k\tau C_{A0}}}{2k\tau}$

example

Phenol Production in CSTR

Consider the reaction of cumene hydroperoxide (CHP) to phenol and acetone



$$(C_6H_5)C(CH_3)_2OOH \longrightarrow (C_6H_5)OH + (CH_3)_2CO$$

$$\text{CHP} \longrightarrow \text{phenol} + \text{acetone}$$

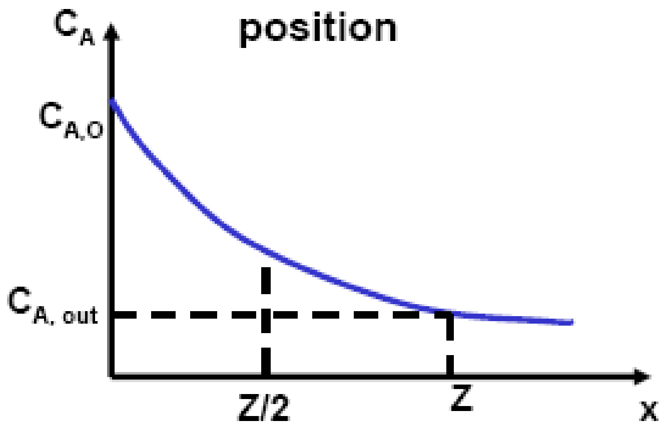
The reaction is pseudo-first-order

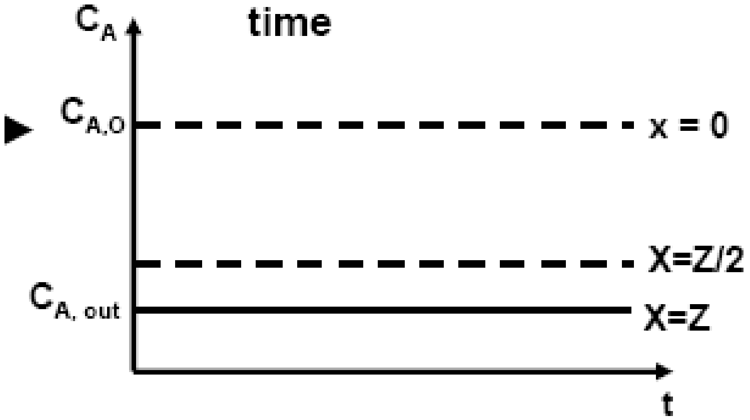
$$r = kc_{\text{CHP}}$$

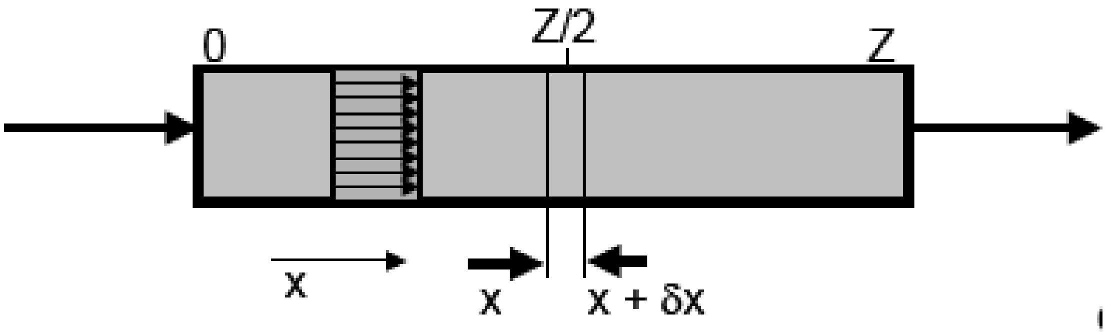
Find the reactor volume to achieve 85% conversion of CHP at steady state. The flowrate into the reactor is $Q_f = 26.9\ \text{m}^3/\text{hr}$ and $k = 4.12\ \text{hr}^{-1}$.

Plug-flow reactor (PFR)

- steady-state
- maintenance of constant throughput
- turbulent flow with plug flow velocity profile
- L/D ratio is sufficiently large
- D/Dp ratio is sufficiently large
- low pressure drop
- well-defined axial T-profile
- negligible radial T-profile
- prevention of further reaction following sampling







Plug-flow reactor (PFR)

- A plug-flow reactor (PFR) may be used for both liquid-phase and gas-phase reactions, and for both laboratory-scale investigations of kinetics and large-scale production. The reactor itself may consist of an empty tube or vessel, or it may contain packing or a tied bed of particles (e.g., catalyst particles).
- A PFR is similar to a CSTR in being a flow reactor, but is different in its mixing characteristics. It is different from a BR in being a flow reactor, but is similar in the progressive change of properties, with position replacing time. These features are explored further in this section, but first we elaborate the characteristics of a PFR, as follows:

1. The flow through the vessel, both input and output streams, is continuous, but not necessarily at constant rate; the flow in the vessel is PF.
2. The system mass inside the vessel is not necessarily fixed.
3. There is no axial mixing of fluid inside the vessel (i.e., in the direction of flow).

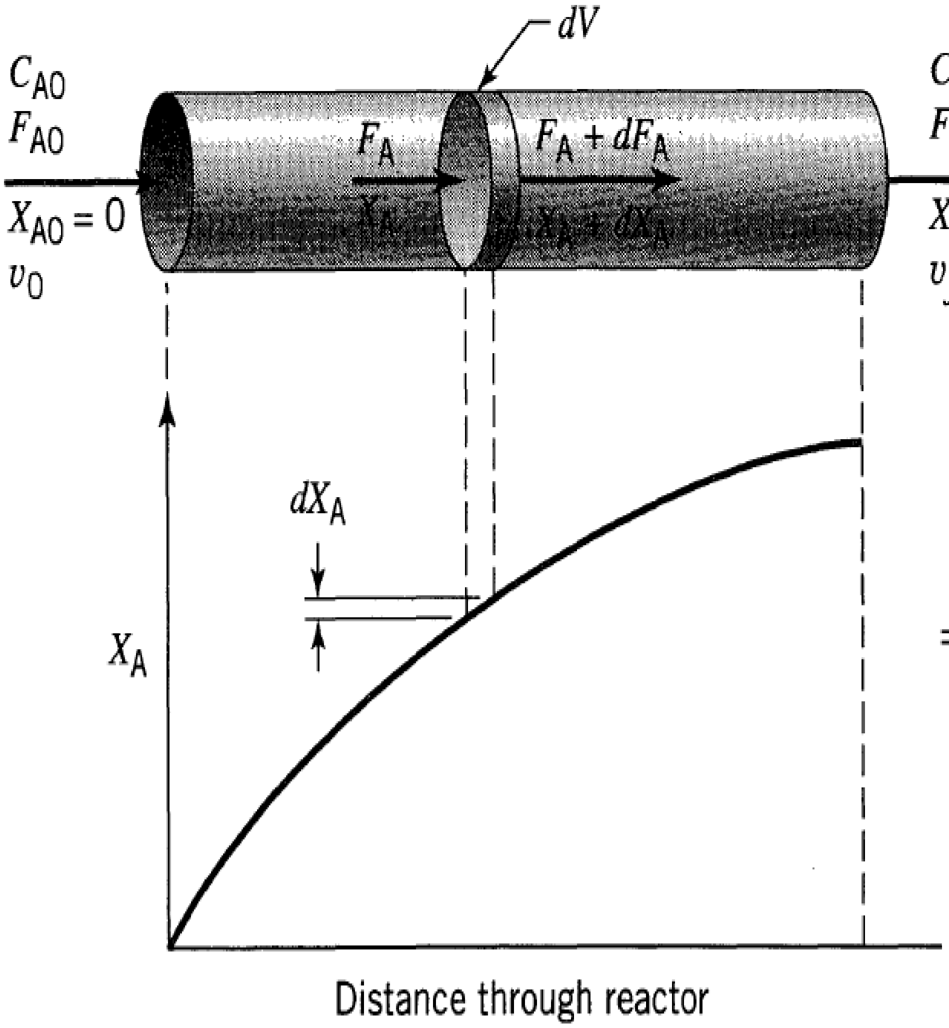
Continued…..

4. There is complete radial mixing of fluid inside the vessel (i.e., in the plane perpendicular to the direction of flow); thus, the properties of the fluid, including its velocity, are uniform in this plane.
5. The density of the flowing system may vary in the direction of flow.
6. The system may operate at steady-state or at unsteady-state.
7. There may be heat transfer through the walls of the vessel between the system and the surroundings.

- **MATERIAL BALANCE:**
- In a plug flow reactor the composition of the fluid varies from point to point along a flow path; consequently, the material balance for a reaction component must be made for a differential element of volume dV. Thus for reactant **A,** the balance equation becomes

$$\text{input} = \text{output} + \text{disappearance by reaction} + \overset{=0}{\cancel{\text{accumulation}}}$$

Performance equation of PFR



Notation for a plug flow reactor.

from the figure

$$\text{input of A, moles/time} = F_A$$

$$\text{output of A, moles/time} = F_A + dF_A$$

$$\text{disappearance of A by reaction, moles/time} = (-r_A)dV$$

$$= \left(\frac{\text{moles A reacting)}}{\text{(time)(volume of fluid)}}\right)\left(\begin{matrix}\text{volume of}\\ \text{element}\end{matrix}\right)$$

Therefore the overall balance equation is as follows:

$$F_A = (F_A + dF_A) + (-r_A)dV$$

or

$$dF_A = d[F_{A0}(1 - X_A)] = -F_{A0}dX_A$$

or

$$F_{A0}dX_A = (-r_A)dV$$

Continued…

- This, then, is the equation which accounts for A in the differential section of reactor of volume dV. For the reactor as a whole the expression must be integrated. Now FA,, the feed rate, is constant, but r, is certainly dependent on the concentration or conversion of materials. Grouping the terms accordingly, we obtain

$$\int_0^V \frac{dV}{F_{A0}} = \int_0^{X_{Af}} \frac{dX_A}{-r_A}$$

- Thus

$$\boxed{\begin{aligned} \frac{V}{F_{A0}} &= \frac{\tau}{C_{A0}} = \int_0^{X_{Af}} \frac{dX_A}{-r_A} \\ \tau = \frac{V}{v_0} &= \frac{VC_{A0}}{F_{A0}} = C_{A0}\int_0^{X_{Af}} \frac{dX_A}{-r_A} \end{aligned}} \qquad \text{any } \varepsilon_A$$

Continued…

SOME SPECIAL CASES

- For constant-density systems ($\varepsilon_A = 0$)

$$\frac{V}{F_{A0}} = \frac{\tau}{C_{A0}} = \int_0^{X_{Af}} \frac{dX_A}{-r_A} = -\frac{1}{C_{A0}} \int_{C_{A0}}^{C_{Af}} \frac{dC_A}{-r_A}$$

$$\tau = \frac{V}{v_0} = C_{A0} \int_0^{X_{Af}} \frac{dX_A}{-r_A} = -\int_{C_{A0}}^{C_{Af}} \frac{dC_A}{-r_A}$$

- Zero-order homogeneous reaction, any constant ε_A

$$k\tau = \frac{kC_{A0}V}{F_{A0}} = C_{A0}X_A$$

- First-order irreversible reaction

$$A \rightarrow \text{products, any constant } \varepsilon_A,$$

$$k\tau = -(1+\varepsilon_A)\ln(1-X_A) - \varepsilon_A X_A$$

Continued…

- First-order reversible reaction,

$A \rightleftarrows rR,\ C_{R0}/C_{A0} = M,$ and $-r_A = k_1 C_A - k_2 C_R$

therefore the performance equation is

$$k_1\tau = \frac{M + rX_{Ae}}{M + r}\left[-(1 + \varepsilon_A X_{Ae})\ln\left(1 - \frac{X_A}{X_{Ae}}\right) - \varepsilon_A X_A\right]$$

where X_{Ae} is the equilibrium conversion.

- Second-order irreversible reaction

$2A \rightarrow$ products, any constant ε_A

And the performance equation is

$$C_{A0}k\tau = 2\varepsilon_A(1 + \varepsilon_A)\ln(1 - X_A) + \varepsilon_A^2 X_A + (\varepsilon_A + 1)^2 \frac{X_A}{1 - X_A}$$

Holding Time and Space Time for Flow Reactors

- Space time is

$$\tau = \begin{pmatrix} \text{time needed to} \\ \text{treat one reactor} \\ \text{volume of feed} \end{pmatrix} = \frac{V}{v_0} = \frac{C_{A0}V}{F_{A0}}, \quad [\mathrm{hr}]$$

- Holding time is

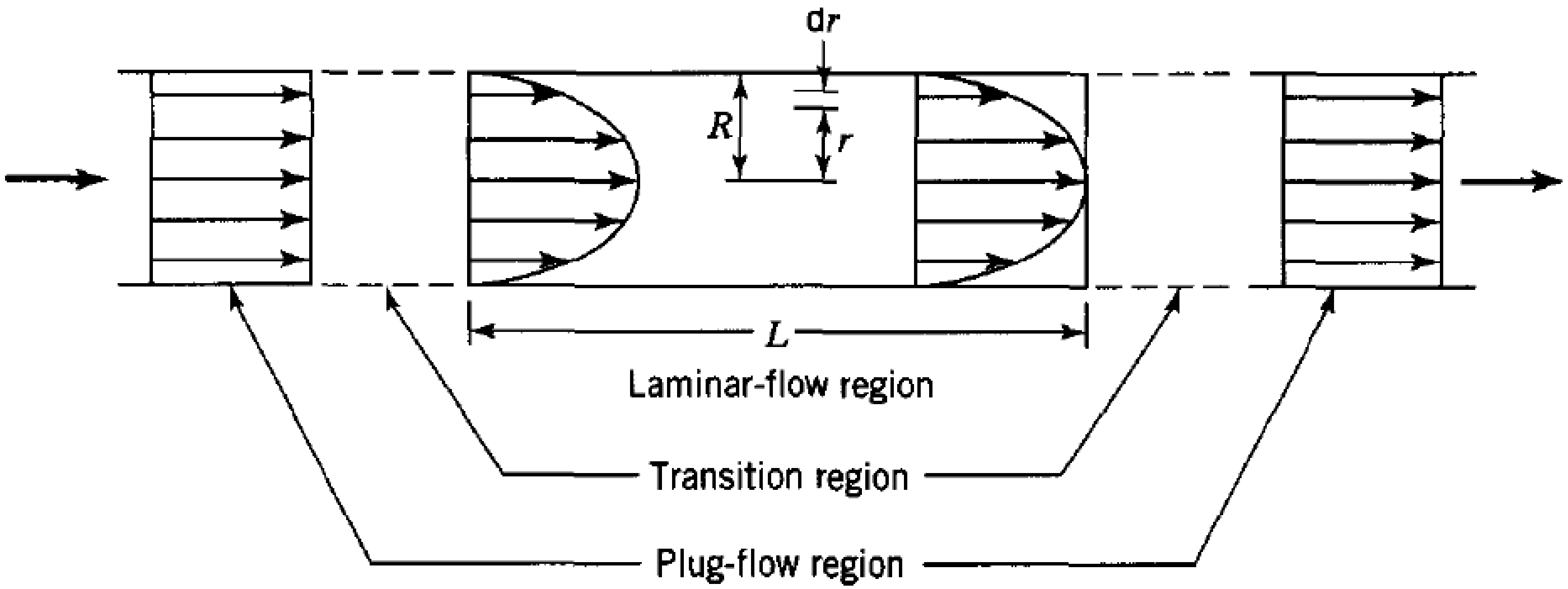
$$\bar{t} = \begin{pmatrix} \text{mean residence time} \\ \text{of flowing material} \\ \text{in the reactor} \end{pmatrix} = C_{A0}\int_0^{X_A} \frac{dX_A}{(-r_A)(1+\varepsilon_A X_A)}, \quad [\mathrm{hr}]$$

- Therefore for constant density systems

$$\tau = \bar{t} = \frac{V}{v}$$

Laminar-flow reactor (LFR)

- A laminar-flow reactor (LFR) is rarely used for kinetic studies, since it involves a flow pattern that is relatively difficult to attain experimentally. However, the *model* based on laminar flow, a type of tubular flow, may be useful in certain situations, both in the laboratory and on a large scale, in which flow approaches this extreme (at low Re). Such a situation would involve low fluid flow rate, small tube size, and high fluid viscosity, either separately or in combination, as, for example, in the extrusion of high-molecular-weight polymers.



LAMINAR FLOW REACTOR

Characteristics of LFR

- The general characteristics of the simplest model of a continuous LFR, illustrated schematically in above Figure , are as follows:

1. The flow through the vessel is laminar (LF) and continuous, but not necessarily at constant rate.
2. The system mass inside the vessel is not necessarily fixed.
3. There is no axial mixing of fluid inside the vessel.
4. There is no radial mixing of fluid inside the vessel.
5. The density of the flowing system is not necessarily constant.
6. The system may operate at steady-state or at unsteady-state.
7. There may be heat transfer through the walls of the vessel between the system and the surroundings.

Size comparison of single reactors

- The batch reactor has the advantage of small instrumentation cost and flexibility of operation (may be shut down easily and quickly).
- It has the disadvantage of high labor and handling cost, often considerable shutdown time to empty, clean out, and refill, and poorer quality control of the product.
- Hence we may generalize to state that the batch reactor is well suited to produce small amounts of material and to produce many different products from one piece of equipment.
- On the other hand, for the chemical treatment of materials in large amounts the continuous process is nearly always found to be more economical.
- Regarding reactor sizes, a comparison for a given duty and for $\varepsilon = 0$ shows that an element of fluid reacts for the same length of time in the batch and in the plug flow reactor have the same volume.
- Of course, on a long-term production basis we must correct the size requirement estimate to account for the shutdown time between batches. Still, it is easy to relate the performance capabilities of the batch reactor with the plug flow reactor.

Continued…

- **Mixed Versus Plug Flow Reactors**
- For a given duty the ratio of sizes of mixed and plug flow reactors will depend on the extent of reaction, the stoichiometry, and the form of the rate equation.
- Let us make this comparison for the large class of reactions approximated by the simple nth-order rate law

$$-r_A = -\frac{1}{V}\frac{dN_A}{dt} = kC_A^n$$

- For mixed flow reactor the space time calculated is as follows:

$$\tau_m = \left(\frac{C_{A0}V}{F_{A0}}\right)_m = \frac{C_{A0}X_A}{-r_A} = \frac{1}{kC_{A0}^{n-1}}\frac{X_A(1+\varepsilon_A X_A)^n}{(1-X_A)^n}$$

- For plug flow reactor the space time calculated is as follows:

$$\tau_p = \left(\frac{C_{A0}V}{F_{A0}}\right)_p = C_{A0}\int_0^{X_A}\frac{dX_A}{-r_A} = \frac{1}{kC_{A0}^{n-1}}\int_0^{X_A}\frac{(1+\varepsilon_A X_A)^n dX_A}{(1-X_A)^n}$$

Continued…

- Therefore

$$\frac{(\tau C_{A0}^{n-1})_m}{(\tau C_{A0}^{n-1})_p}=\frac{\left(\dfrac{C_{A0}^{n}V}{F_{A0}}\right)_m}{\left(\dfrac{C_{A0}^{n}V}{F_{A0}}\right)_p}=\frac{\left[X_A\left(\dfrac{1+\varepsilon_A X_A}{1-X_A}\right)^n\right]_m}{\left[\displaystyle\int_0^{X_A}\left(\dfrac{1+\varepsilon_A X_A}{1-X_A}\right)^n dX_A\right]_p}$$

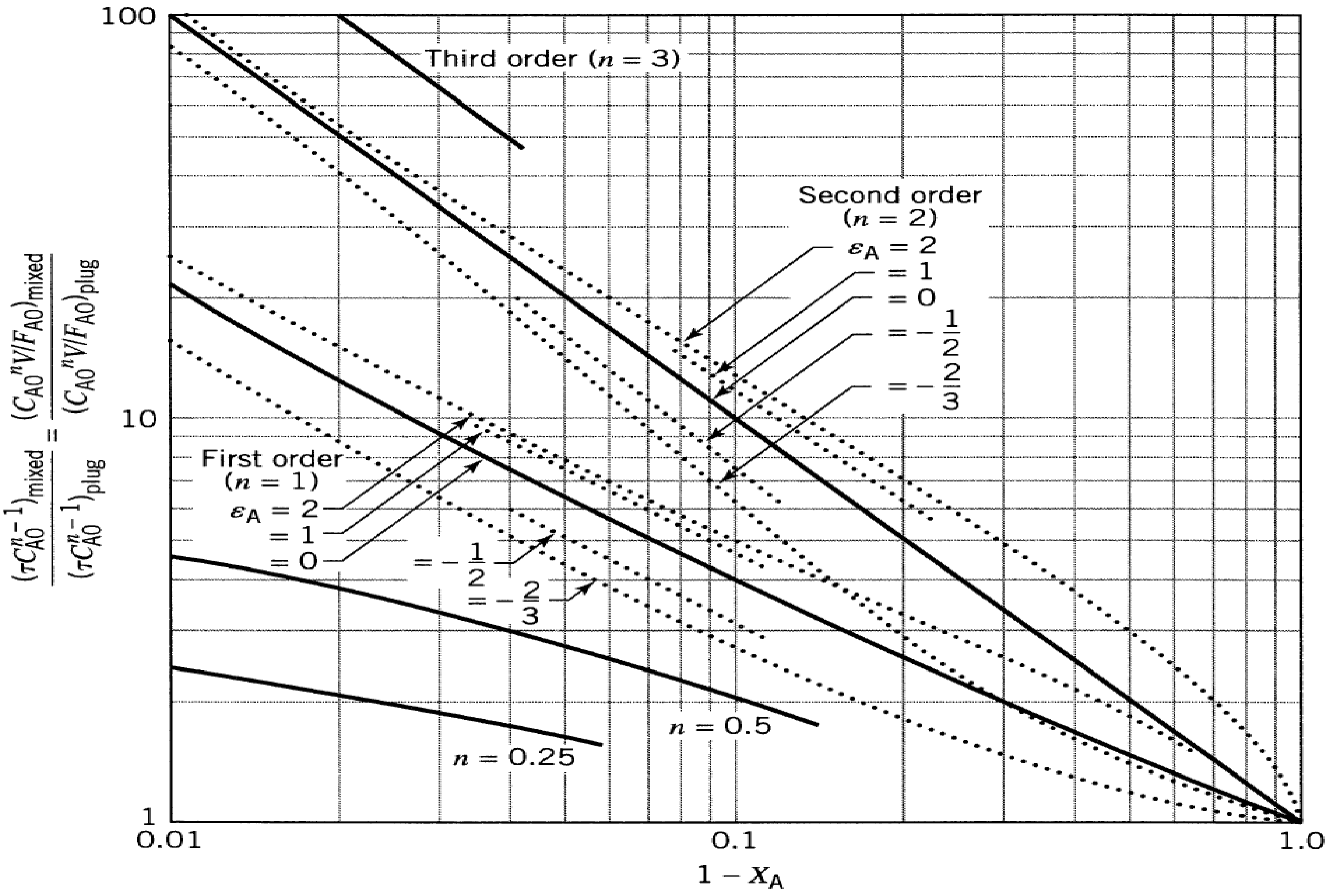
- With constant density, or ε = 0, this expression integrates to

$$\frac{(\tau C_{A0}^{n-1})_m}{(\tau C_{A0}^{n-1})_p}=\frac{\left[\dfrac{X_A}{(1-X_A)^n}\right]_m}{\left[\dfrac{(1-X_A)^{1-n}-1}{n-1}\right]_p},\qquad n\neq 1$$

- or

$$\frac{(\tau C_{A0}^{n-1})_m}{(\tau C_{A0}^{n-1})_p}=\frac{\left(\dfrac{X_A}{1-X_A}\right)_m}{-\ln(1-X_A)_p},\qquad n=1$$

Comparison of performance of single mixed flow and plug flow reactors for the nth-order reactions



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- From the above figure we can conclude that:

1. For any particular duty and for all positive reaction orders the mixed reactor is always larger than the plug flow reactor. The ratio of volumes increases with reaction order.
2. When conversion is small, the reactor performance is only slightly affected by flow type. The performance ratio increases very rapidly at high conversion; consequently, a proper representation of the flow becomes very important in this range of conversion.
3. Density variation during reaction affects design; however, it is normally of secondary importance compared to the difference in flow type.

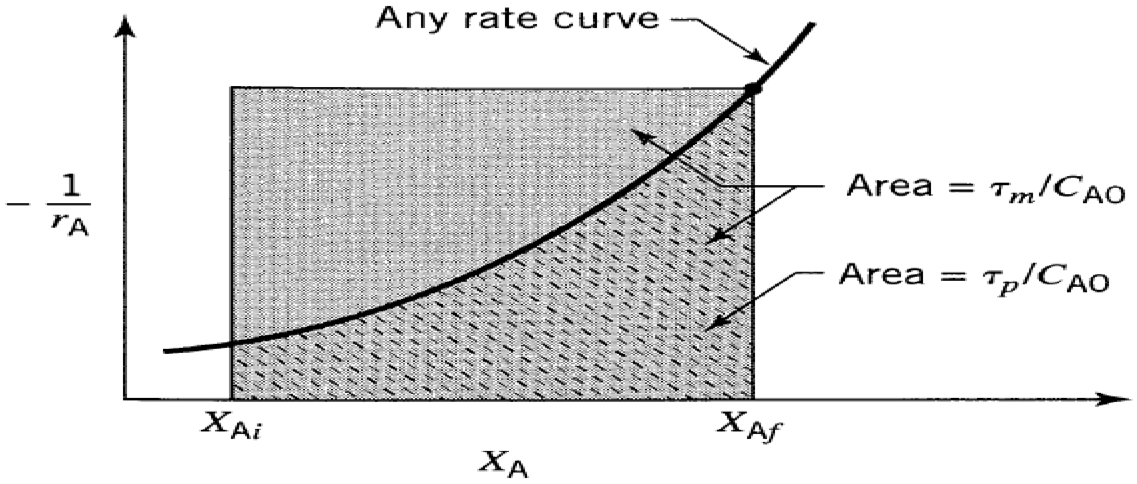
- **Variation of Reactant Ratio for Second-Order Reactions**
- Second-order reactions of two components and of the type

$$A + B \rightarrow \text{products}, \qquad M = C_{B0}/C_{A0}$$

$$-r_A = -r_B = kC_AC_B$$

- For ***M* =*1***, the order of reaction is two and for ***M* > *1***, the order of reaction is one.

General Graphical Comparison



Comparison of performance of mixed flow and plug flow reactors for any reaction kinetics.

- For reactions with arbitrary but known rate the performance capabilities of mixed and plug flow reactors are best illustrated in Fig. The ratio of shaded and of hatched areas gives the ratio of space-times needed in these two reactors.
- The rate curve drawn in Fig. 6.2 is typical of the large class of reactions whose rate decreases continually on approach to equilibrium (this includes all nth-order reactions, n > 0). For such reactions it can be seen that mixed flow always needs a larger volume than does plug flow for any given duty.

Multiple-reactor systems

- **Plug Flow Reactors in Series and/or in Parallel**
- Consider N plug flow reactors connected in series, and let $X_1, X_2, \ldots, X_N$, be the fractional conversion of component A leaving reactor 1, 2, . . . , N. Basing the material balance on the feed rate of A to the first reactor, we find for the ith reactor from

$$\frac{V_i}{F_0} = \int_{X_{i-1}}^{X_i} \frac{dX}{-r}$$

- For N numbers of reactors in series

$$\frac{V}{F_0} = \sum_{i=1}^{N} \frac{V_i}{F_0} = \frac{V_1 + V_2 + \cdots + V_N}{F_0}$$

$$= \int_{X_0=0}^{X_1} \frac{dX}{-r} + \int_{X_1}^{X_2} \frac{dX}{-r} + \cdots + \int_{X_{N-1}}^{X_N} \frac{dX}{-r} = \int_0^{X_N} \frac{dX}{-r}$$

- Hence, N plug flow reactors in series with a total volume V gives the same conversion as a single plug flow reactor of volume V.

Continued…

- For the optimum hook up of plug flow reactors connected in parallel or in any parallel-series combination, we can treat the whole system as a single plug flow reactor of volume equal to the total volume of the individual units if the feed is distributed in such a manner that fluid streams that meet have the same composition. Thus, for reactors in parallel V/F or τ must be the same for each parallel line. Any other way of feeding is less efficient.

Equal-Size Mixed Flow Reactors in Series

- In plug flow, the concentration of reactant decreases progressively through the system; in mixed flow, the concentration drops immediately to a low value. Because of this fact, a plug flow reactor is more efficient than a mixed flow reactor for reactions whose rates increase with reactant concentration, such as nth-order irreversible reactions, $n > 0$.
- Consider a system of N mixed flow reactors connected in series. Though the concentration is uniform in each reactor, there is, nevertheless, a change in concentration as fluid moves from reactor to reactor. This stepwise drop in concentration, illustrated in Fig. below, suggests that the larger the number of units in series, the closer should the behavior of the system approach plug flow.

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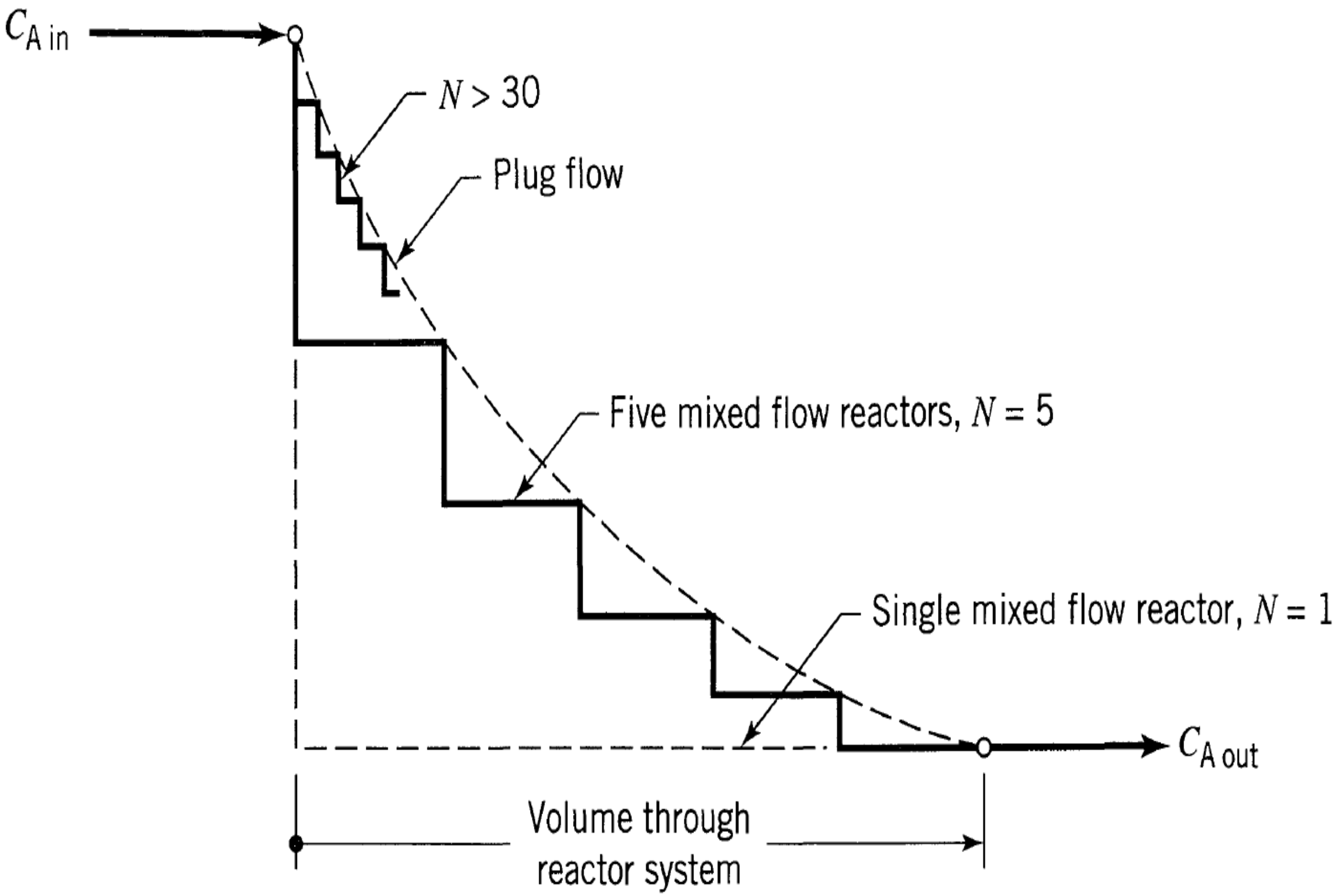

Concentration profile through an N-stage mixed flow reactor system compared with single flow reactors.

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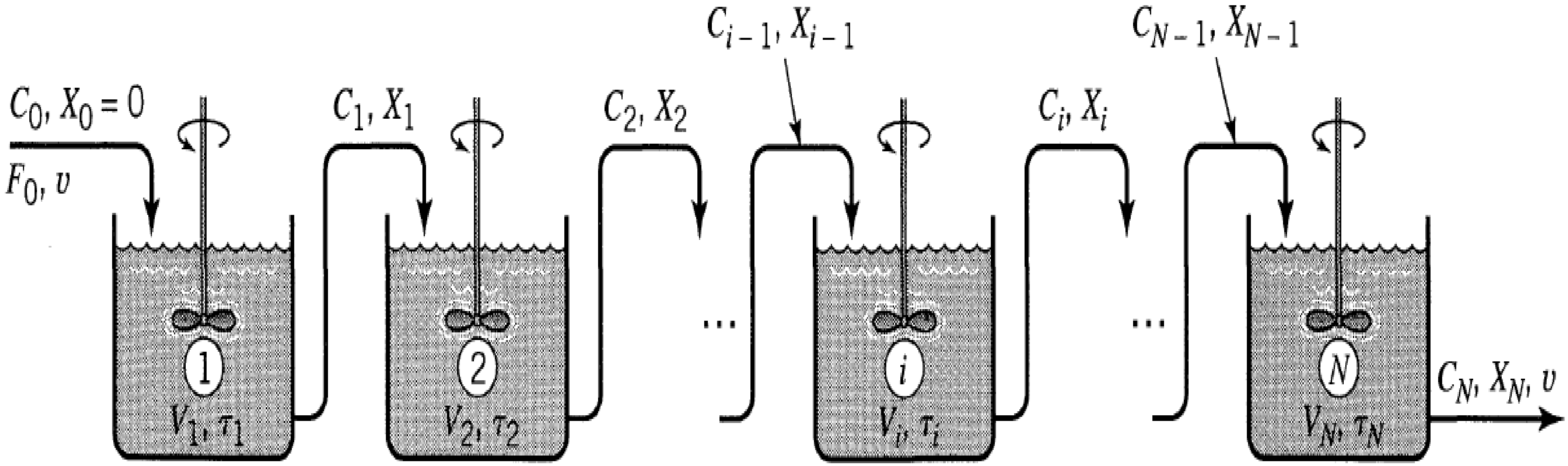


Notation for a system of N equal-size mixed reactors in series.

- Let us now quantitatively evaluate the behavior of a series of N equal-size mixed flow reactors. Density changes will be assumed to be negligible; hence ε = 0 and t = τ . As a rule, with mixed flow reactors it is more convenient to develop the necessary equations in terms of concentrations rather than fractional conversions; therefore, we use this approach. The nomenclature used is shown in Fig. above with subscript i referring to the ith vessel.

Continued…

- So the performance equation becomes

$$\tau_i = \frac{C_0 V_i}{F_0} = \frac{V_i}{v} = \frac{C_0(X_i - X_{i-1})}{-r_{Ai}}$$

- For 1st order reaction in constant density system

$$\tau_i = \frac{C_0[(1 - C_i/C_0) - (1 - C_{i-1}/C_0)]}{kC_i} = \frac{C_{i-1} - C_i}{kC_i}$$

Or

$$\frac{C_{i-1}}{C_i} = 1 + k\tau_i$$

- Now the space-time (or mean residence time t) is the same in all the equal size reactors of volume V_i. Therefore,

$$\frac{C_0}{C_N} = \frac{1}{1 - X_N} = \frac{C_0}{C_1}\frac{C_1}{C_2}\cdots\frac{C_{N-1}}{C_N} = (1 + k\tau_i)^N$$

Continued…

- Rearranging, we find for the system as a whole

$$\tau_{N\,\mathrm{reactors}} = N\tau_i = \frac{N}{k}\left[\left(\frac{C_0}{C_N}\right)^{1/N} - 1\right]$$

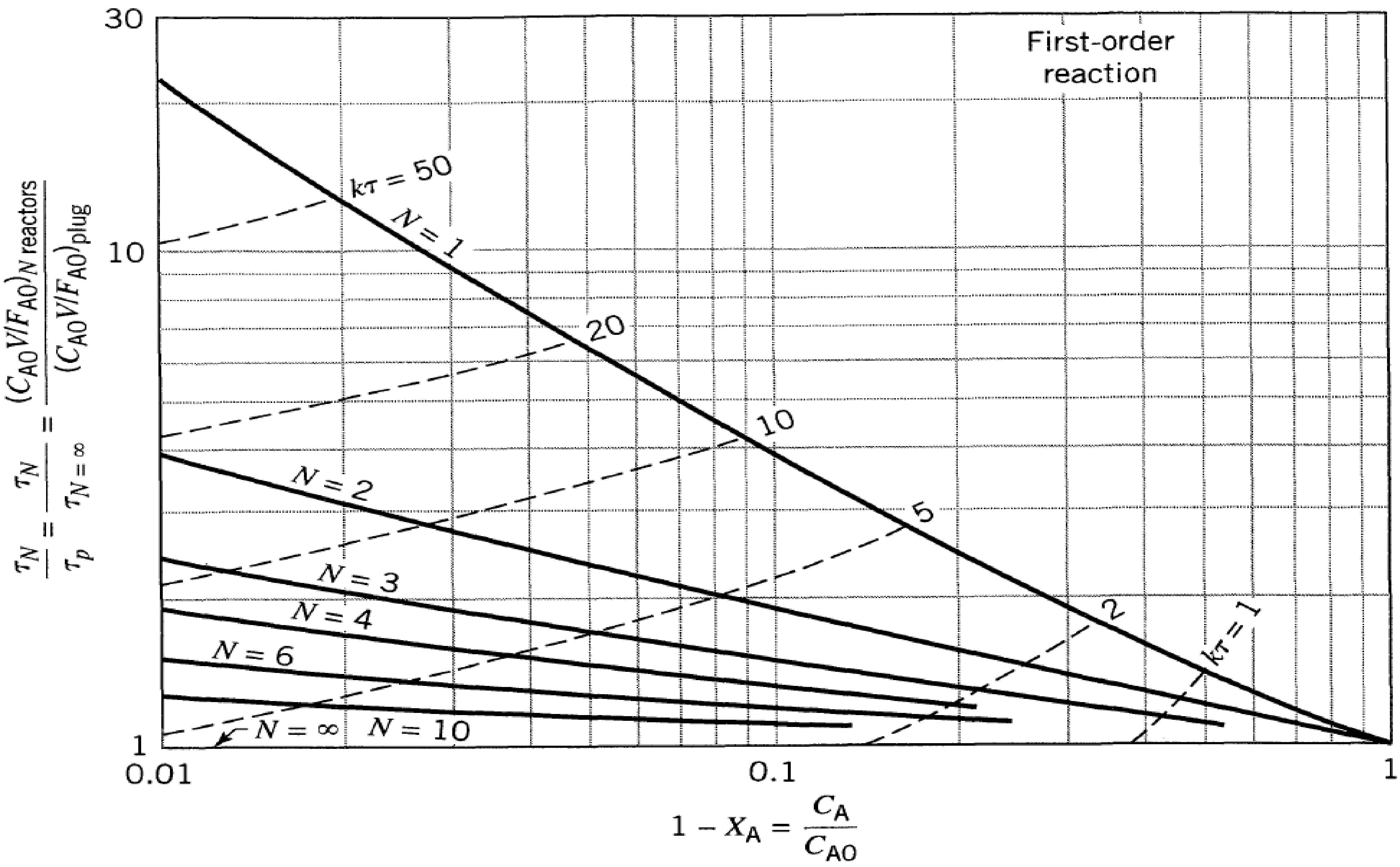
- In the limit, for N →∞ , this equation reduces to the plug flow equation

$$\tau_p = \frac{1}{k}\ln\frac{C_0}{C}$$

- This comparison is shown in Fig. below for first-order reactions in which density variations are negligible.

Continued…

- **Comparison of performance of a series of N equal-size mixed flow reactors with a plug flow reactor for the first-order reaction**



PFR and CSTR Combinations in Series

- To this point it has been shown, mainly by illustration, that the performance of a reactor depends on three factors:

1. The kinetics of reaction, as described by the rate law; thus, the fractional conversion depends on the order of reaction.
2. The characteristics of flow through the vessel as described by the RTD; thus, for given kinetics, the fractional conversion in a reactor with PF (a PFR) is different from that in one with BMF (a CSTR).
3. The nature of mixing of fluid elements during flow through the vessel as characterized, so far, by the degree of segregation; thus, in a CSTR (i.e., a given RTD) with given kinetics, the fractional conversion is different for the extremes of non-segregated flow and completely segregated flow, except for a first-order reaction, which is a linear process. Reactions of other orders are examples of nonlinear processes