School of Engineering
Chemical Engineering Department

(0905421) Chemical Reaction Engineering I — In-Class Assessment # 3 (Chapter 1)

Partner (1) Name: رغد أحمد السعدي
Partner (2) Name: [illegible]

Identity (1) #: 0153646
Identity (2) #: 0155794

Self-Test – What Have You Learned?

1. Which equation is used in arriving at the design equation for a batch reactor?
a. $G_j = V*r_j$
b. $dN_j/dt = 0$
[✓ c. $F_{j0} = F_j = 0$]
d. $E = mc^2$

2. What assumptions are made when modeling an ideal tubular reactor? ? PFR
[✓ a. Steady sate and no radial variations]
b. Plug flow and liquid systems
c. Gas flow and steady state
d. That the reactor will photograph well

3. What does the mole balance for a CSTR become if $r_j = -k\,C_j$?
a. $C_j = (F_{j0}-F_j)/V$
b. too complicated
[✓ c. $V = (F_{j0} - F_j)/(k*C_j)$]
d. $V = (F_{j0} - F_j)/k$

4. What is the name for any chemical compound or element with a given identity?
a. Ion
b. Atom
c. Alfred
[✓ d. Chemical species]

5. Which of the following is a combination type reaction?
a. 53 left: 37 right: 5 left
[✓ b. $A + B \rightarrow C$]
c. $A_2 \rightarrow 2A$
d. $A \rightarrow B$

6. What happens during a combination reaction?
a. A molecule breaks down into smaller molecules
b. You get a burger, fries and a soft drink
[✓ c. Two or molecules combine]
d. Species' molecular configuration changes only

7. What happens during a decomposition reaction?
a. Species' molecular configuration changes only
b. A molecule rots
c. Two molecules combine to give one molecule
[✓ d. A molecule breaks down into smaller molecules]

8. What assumption is made when studying an IDEAL CSTR?
a. Innocent until proven guilty
b. Adiabatic operation
[✓ c. Perfect mixing]
d. Constant volume

9. Which of the following reactors usually have the easiest temperature control?
a. Blue reactors
[✓ b. CSTRs]
const. temp. at all rxn
c. Tubular reactors
d. Batch reactors

10. A packed bed reactor is also known as a fluidized bed reactor. ⇒ CSTR مسمى ثاني
a. Only when using an alias
b. Depends on the catalyst used
c. True
[✓ d. Always false]

11. What type of mathematical equation is used to express the rate law?
a. Irreversible equation
b. Differential equation
c. Unsolvable equation
[✓ d. Algebraic equation]

12. What are the dimensions of k in the equation $-r_A = k\,C_A$?
a. $moles^2/volume^2/time$
b. time/mole/volume
c. three
[✓ d. 1/time]

What is Wrong with This Solution?

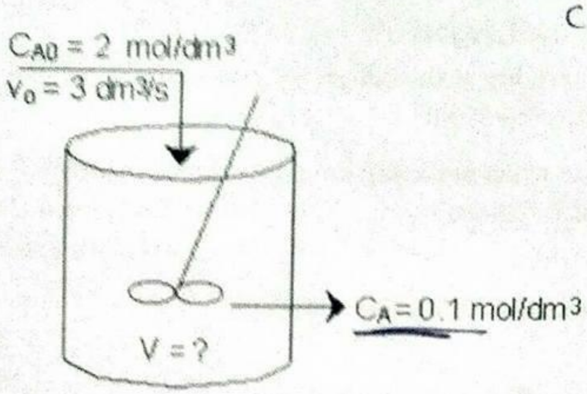
Problem

The irreversible liquid phase second order reaction $2A \xrightarrow{k_1} B$ $(k_1 = 0.03\ dm^3/mol{\bullet}s)$ is carried out in a CSTR. The entering concentration of A, C_{A0}, is 2 M and the exit concentration of A, C_A is 0.1 M. The entering and exiting volumetric flow rate, υ_0, is constant at 3 dm³/s. What is the corresponding reactor volume?

Solution

$C_{Ao} = 2\ M$

$C_{Aout} = 0.1\ M$



Mole Balance $$V = \frac{F_{A0} - F_A}{-r_A}$$

Rate Law $$-r_A = kC_A^2\ (2^{nd}\ \text{order})$$

Combine $$V = \frac{F_{A0} - F_A}{kC_A^2}$$

$$F_{A0} = v_o C_{A0} = \frac{3\ dm^3}{s} \bullet \frac{2\ molA}{dm^3} = \frac{6\ molA}{s}$$

$$F_A = v_o C_A = \frac{3\ dm^3}{s} \bullet \frac{0.1\ molA}{dm^3} = \frac{0.3\ molA}{s}$$

$$V = \frac{(6-0.3)\frac{mol}{s}}{\left(0.03\frac{dm^3}{mol\bullet s}\right)\left(2\frac{mol}{dm^3}\right)^2} = \frac{5.7dm^3}{0.12} = 47.5\ dm^3$$

$C_A = 0.1\ mol/dm^3$ exist

* نستعمل C_A outlet

What is wrong with this solution?

$$\boxed{V = 19{,}000\ dm^3}$$

⇏ حجم كبير

لأنه الـ rate

كثير قليل ← الـ outlet

لوكان الـ rate

الـ input

بكون عالي

بيعطي حجم صغير

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(0905421) Chemical Reaction Engineering I **In-Class Assessment # 2 (Chapter 1)**

Partner (1) Name: / غداء، البيدي **Partner (2) Name:** علا عزت، الحباك

Identity (1) #: 0153646 **Identity (2) #:** 0155794

Self-Test – What Have You Learned?

→ $-r_A = kC_A$ or $-r_A = kC_A^2$

1. The rate law
   - (a) is a differential equation T [✓ F]
   - (b) relates reaction rate and concentration of reacting species [✓ T] F
   - (c) is only valid in the U.S. T [✓ F]
   - (d) is a function of temperature; [✓ T] F

     ↳ and press., conc., catalyst if any
2. The rate of formation of species A is per unit volume is
   - (a) r_A [✓ T] F
   - (b) $-r_A$ T [✓ F]
   - (c) the rate of generation of species A per unit volume [✓ T] F
   - (d) r'_A T [✓ F]
   - (e) none of the above T [✓ F]
3. Consider the reaction

$$\overset{10 \text{ disappearance}}{\leftarrow} A \rightarrow B \overset{10 \text{ formation}}{\rightarrow}$$

At a particular time, t, the rate of formation of B in the reaction, $\boxed{r_B\text{, is } 10\ \text{mole/dm}^3\bullet\text{min}}$. Which of the following are true?

- (a) The rate of disappearance of B is -10 moles/dm^3*min. T
- (b) The rate of formation of A is -10 mole/dm^3*min. T
- (c) The rate of disappearance of A is 10 moles/dm^3*min. T
- (d) $r_A = -10$ moles/dm^3*min T
- (e) $-r_A =$ 10 moles/dm^3*min T
- (f) $-r_B = -10$ moles/dm^3*min T
- (g) Some of the above
- [✓ (h)] All of the above
- (i) None of the above

Q2: For the reaction

$$A \rightarrow B$$

Calculate the time to reduce the number of moles by a factor of 10 ($N_A = \frac{N_{A0}}{10}$) in a batch reactor for the above reaction with $-r_A = kC_A$, when $k = 0.046\ \text{min}^{-1}$ $N_A = 0.1 N_{A0}$

$$t = \int_{N_{A0}}^{N_A} \frac{dN_A}{r_A V}$$

$$= \int_{N_{A0}}^{N_{A0}/10} \frac{dN_A}{-k C_A V} \quad ; \boxed{N_A = C_A V}$$

$$= \frac{+1}{k} \int_{N_{A0}/10}^{N_{A0}} \frac{dN_a}{N_a}$$

$$= \frac{1}{k} \left(\ln N_{A0} - \ln \frac{N_A}{10} \right)$$

$$= \frac{1}{k} \ln \left(\frac{N_{A0}}{\frac{N_{A0}}{10}} \right)$$

$$= \frac{1}{k} \ln (10)$$

$$\boxed{t = 50\ \text{min}}$$

$$\cancel{\frac{\Delta N_A}{r_A V}} = \cancel{\frac{N_A - N_{A0}}{r_A V}}$$

$$\Rightarrow \cancel{\frac{0.1 N_{A0} - N_{A0}}{r_A V}}$$

$$\cancel{t} = \cancel{\frac{-0.9 N_{A0}}{r_A V}}$$

$$\cancel{t} = \cancel{\frac{-0.9 C_A}{-k C_A}}$$

$$= \cancel{\frac{0.9}{0.046}}$$

$$= \cancel{19.56}$$

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(0905421) Chemical Reaction Engineering I In-Class Assessment # 1 (Chapter 1)

Partner (1) Name: [illegible] **Partner (2) Name:** [illegible]

Identity (1) #: 0153646 **Identity (2) #:** 0155794

Q1: Consider the reaction

$$A + 2B \rightarrow 3C$$

5

in which the rate of disappearance of A is 5 moles of A per dm^3 per second at the start of the reaction.At the start of the reaction:

(a) What is $-r_A$? 5 mol/dm^3.s

(b) What is the rate of formation of B?

$r_B = -10$ mol/dm^3.s

(c) What is the rate of formation of C?

$r_c = +15$ mol/dm^3.s

(d) What is the rate of disappearance of C?

$-r_c = -15$ mol/dm^3.S

(e) What is the rate of formation of A, r_A?

$r_A = -5$ mol/dm^3.s

(f) What is $-r_B$? rate of disappearent.

$-r_B = 10$ mol/dm^3.s

Levenspiel Plots in Terms of ~~Conversion~~ concentration

For reactions in which the rate depends only on the concentration of one species [i.e., $-r_A = f(C_A)$], it is usually convenient to report $-r_A$ as a function of concentration rather than conversion. We can rewrite the design equation for a plug-flow reactor PFR [Equation (1)] in terms of the concentration, C_A, rather than in terms of conversion for the special case when $\boxed{\upsilon = \upsilon_0.}$ liq phase

$$V = F_{A0} \int_0^X \frac{dX}{-r_A} \qquad \frac{F_{A0}}{-r_A} X = V \qquad (1)$$

$$F_{A0} = \upsilon_0 C_{A0} \qquad (2)$$

Now write X in terms of entering and exiting molar flow rate, F_{A0} and F_A :

$$\boxed{X = \cancel{\frac{V * -r_A}{F_A + F_{A0}(X)}} = \frac{F_{A0} - F_A}{F_{A0}}} \qquad (3)$$

$F_{A0} - F_A \rightarrow$ reacted

$F_{A0} \rightarrow$ fed

For the special case when $\upsilon = \upsilon_0$, plugging Equation (2) into Equation (3), we get

$$\boxed{X = \cancel{\frac{V * -r_A}{F_A + V C_{A0}(X)}} = \frac{C_{A0} V_0 - C_A \overset{= V_0}{V}}{C_{A0} V_0}} \qquad (4)$$

$$\boxed{= \frac{C_{A0} - C_A}{C_{A0}}}$$

when $X = 0$, $C_A = C_{A0}$

when $X = X$, $C_A = C_A$

Differentiating Equation (4) yields

$C_{A0} X = C_{A0} - C_A$

$C_{A0} dx = 0 - dC_A$

$$dX = \frac{0 - dC_A}{C_{A0}} = -\frac{dC_A}{C_{A0}} \qquad (5)$$

$\Rightarrow$ $dx = -dC_A / C_{A0}$

And substituting in Equation (1):

~~$f_A = f_{A0} - f_{A0}(x)$~~

~~$f_{A0} = f_A + f_{A0}(x)$~~

$$V = \frac{F_{A0}}{C_{A0}} \int_{C_A}^{C_{A0}} \pm \frac{dC_A}{C_{A0}(-r_A)} \qquad (6)$$

$$\boxed{V = V_0 \int_{C_A}^{C_{A0}} \frac{dC_A}{-r_A}}$$

Remembering that $\tau = \frac{V}{\upsilon_0}$, Equation (6) can be rewritten as:

$$\tau = \frac{V}{V_0} = \int_{C_A}^{C_{A0}} \frac{dC_A}{-r_A}$$

Equation (7) is a form of the design equation for constant volumetric flow rate υ_0 that may prove more useful in determining the space time or reactor volume for reaction rates that depend only on the concentration of one species.

Figure 1 shows a typical curve of the reciprocal reaction rate as a function of concentration for an isothermal reaction carried out at constant volume. For reaction orders greater than zero, the rate decreases as concentration decreases. The area under the curve gives the space time necessary to reduce the concentration of A from C_{A0} to C_{A1}.

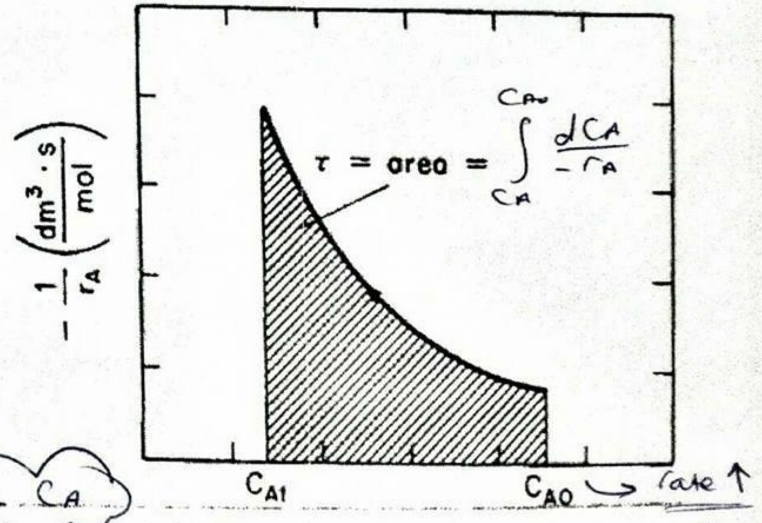

Figure 1 Determining the space time, τ.

$\frac{1}{rate}$ ↓

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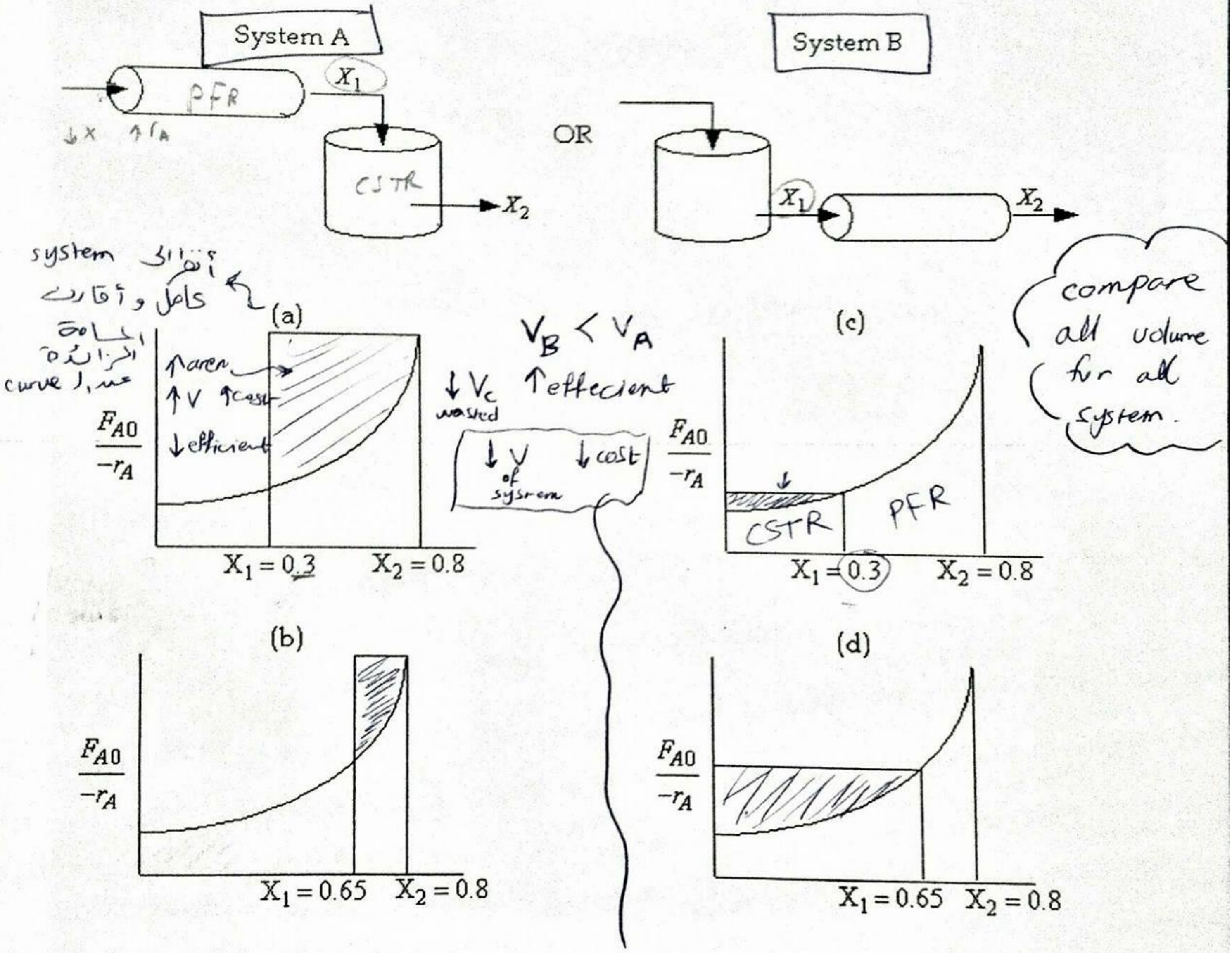
(0905421) Chemical Reaction Engineering I	**In-Class Assessment # 6 (Chapter 2)**
Partner (1) Name: رغد أحمد البدوي	**Partner (2) Name:** علا زكي باكي
Identity (1) #:	**Identity (2) #:**

Self-Test



Q1. Which system is most efficient for an intermediate conversion of (0.3)? B

Q2. Which system is most efficient for an intermediate conversion of (0.65)? A

Q3. Which system makes the best use of the reactor volume (i.e., least "wasted" volume)? Ⓒ

describe C: CSTR flowed by PFR with Intermedial conversion of 0.3.

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(0905421) Chemical Reaction Engineering I	**In-Class Assessment # 4 (Chapter 2)**
Partner (1) Name: رغد البيرقي	**Partner (2) Name:** ملك
Identity (1) #:	**Identity (2) #:**

The gas-phase reaction A + B → C described by the data in the table below is to be carried out at 10 atm and 227 °C. The feed contains 41% A, 41% B, and 18% inert, and flows at a rate of 1 mol/s. F_{A0}

$-r_A$ (mol/(L.min))	0.2	0.0167	0.00488	0.00286	0.00204
X	0.0	0.1	0.4	0.7	0.9

CSTR $X = 0.7$

(a) What is the volume of a single back-mix reactor necessary to achieve 70% conversion?

$$V = \frac{F_{A0} X}{-r_{A\,exist}} = \frac{(1)(0.7)}{0.00286} = 244.755 \quad X \text{ units}$$

$$\frac{1\ \text{mol}}{\text{s}} \bigg| \ 0.7 \ \bigg| \ \frac{\text{L.min}}{0.00286} \bigg| \ 60\text{s} = 14,686\ \text{L}$$

$= 14.700\ m^3$

(b) What is the volume of a single tubular reactor necessary to achieve 60% conversion?

PFR

$X = 0 \rightarrow \frac{F_{A0}}{-r_A} = \frac{1}{+0.2} = 5 \times 60 = 300 = 0.3$ KL

$X = 0.1 \rightarrow \frac{F_{A0}}{-r_A} = \frac{1}{0.0167} \times 60 = 3592.8 = 3.5928$

$X = 0.4 \rightarrow \frac{F_{A0}}{-r_A} = \frac{1}{0.00488} \times 60 = 12295.08 = 12.29508$

$X = 0.6 \rightarrow \frac{F_{A0}}{-r_A} = \frac{1}{3.53 \times 10^{-3}} = 283.3 \times 60 = 18.084$ L

$X = 0.7 \rightarrow \frac{F_{A0}}{-r_A} = \frac{1}{0.00286} \times 60 = 20979 = 20.979$

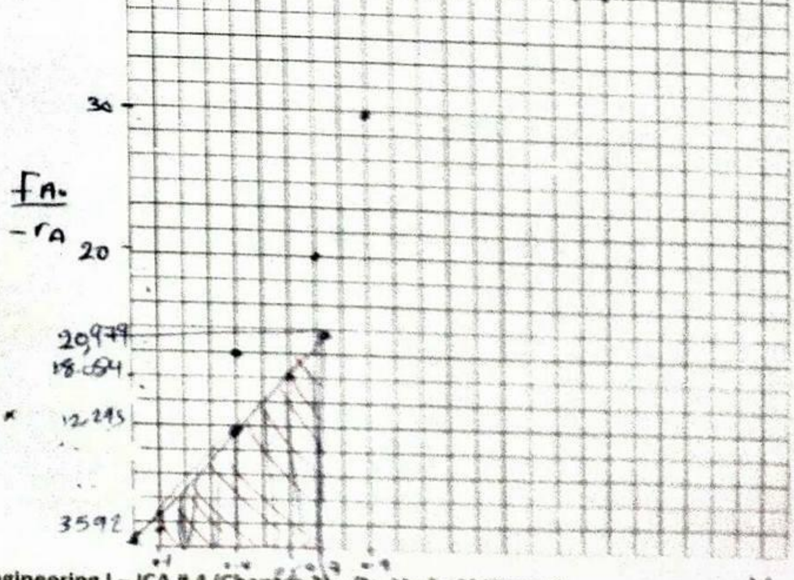
no need $\{ X = 0.9 \rightarrow \frac{F_{A0}}{-r_A} = \frac{1}{0.00204} \times 60 = 29411.76 = 29.41$

= triangle Area

$= \frac{1}{2}(0.6)(18,000)$ (x L)

= 5400 L

= 5.400 m^3



$V = \int_0^X \frac{F_{A0}}{-r_A} dX$ ولكن [illegible] $-r_A$ [illegible] function of (X)

$= \frac{1}{0.00353} \times 60 \times 0.6 =$

at $X = 0.6 \Rightarrow -r_A = 3.53 \times 10^{-3}$ $\frac{\text{mol}}{\text{L/min}}$

$X=0 \Rightarrow V = \frac{F_{A0}}{-r_A} = \frac{1}{0.2} * 60 = 300$ L

$X=0.1 \Rightarrow V = \frac{F_{A0}}{-r_A} = \frac{1}{0.0167} * 60 = 3592.8$ L

$\underline{X=0.4} \Rightarrow V = \frac{F_{A0}}{-r_A} = \frac{1}{0.00488} * 60 = \underline{12,295.08 \text{ L}}$

↓

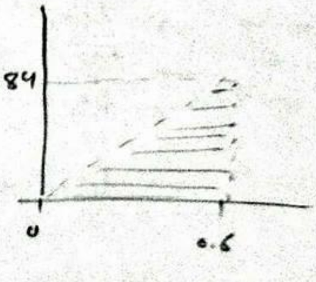
interpolation $\Rightarrow \frac{0.4-0.7}{12,295.08-20,979} = \frac{0.4-0.6}{12295.08-X}$

$X=0.6 \Rightarrow \frac{F_{A0}}{-r_A} = 18,084$ L

↑

$\underline{X=0.7} \Rightarrow V = \frac{F_{A0}}{-r_A} = \frac{1}{0.00286} * 60 = \underline{20,979 \text{ L}}$

Volume in PFR = area under the curve



A under the curve $= V = \frac{1}{2}(x)\left(\frac{F_{A0}}{-r_A}\right)$

$= \left(\frac{1}{2}\right)(0.6)(18,084)$

$= 5425.2$ L

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(0905421) Chemical Reaction Engineering I	**In-Class Assessment # 5 (Chapter 2)**
Partner (1) Name: زين، الدود	**Partner (2) Name:** علا، كيالي
Identity (1) #:	**Identity (2) #:**

$V_0 = 1\ L/s$

A 400-L CSTR and a 100-L PFR are available to process 1.0 L of feed per second. The feed contains 41% A, 41% B, and 18% inert. The irreversible gas-phase reaction A + B → C is to be carried out at 10 atm and 227 °C. The rate of reaction in mol/(L.min) is given below as a function of conversion:

$-r_A$	0.2	0.0167	0.00488	0.00286	0.00204
X	0.0	0.1	0.4	0.7	0.9

in the back (a) What is the maximum conversion that can be achieved with these two reactors connected in series?

$P_0 V_0 = F_{T_0} R T = PV = nRT$

$$C_{A0} = \frac{0.41\ (10\ \text{atm})}{0.08206\ \frac{\text{atm} \cdot L}{\text{mol} \cdot K} * (227+273)\ K} = 0.1\ \text{mol}/L$$

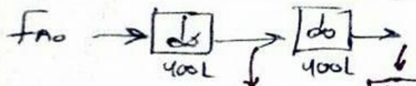
$$\frac{F_{T_0}}{V_0} = C_{T_0} = \frac{P_0}{R T_0}$$

$$C_{A0} = \frac{y_{A0} P_0}{R T_0}$$

$$F_{A0} = 0.1\ \frac{\text{mol}}{L} * 1\ \frac{L}{s} = 0.1\ \text{mol}/s = 6\ \text{mol}/\text{min}$$

$$, X = \frac{V * -r_A}{F_{A0}} =$$

(b) What would be the overall conversion if two 400-L CSTRs were connected in series for the same feed and operating conditions?

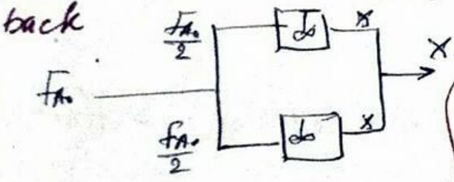


$X_1 = 0.37$ $\boxed{X_2 = 0.59}$

$X_2 = 0.6 \Rightarrow + \frac{F_A}{-r_A} = 1800 \Rightarrow (0.6 - 0.37)(1800) = 414\ L$

$X_2 = 0.5 \Rightarrow \frac{F_A}{-r_A} = 1580 \Rightarrow (0.5 - 0.37)(1580) = 205$

in the back (c) What would be the overall conversion if two 400-L CSTRs were connected in parallel with half of the feed going to each reactor?



$X = 0.52$

$P v = F_{A0} R T$

$\boxed{F_{A0} = C_{A0} v_0}$ $\Rightarrow$ $\boxed{\uparrow \tau = \frac{V}{v \downarrow}}$ $\Rightarrow$ $\boxed{\tau \uparrow \Rightarrow X \uparrow}$

the conversion will be increase

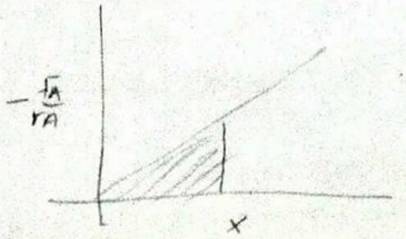
(d) If the total pressure were reduced by a factor of 10, would the conversion increase, decrease, or remain the same? continuous system

$P_2 = 0.1\ P_1$

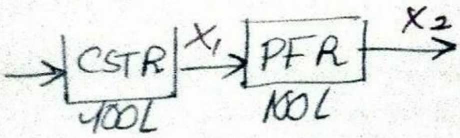
$P v = F R T$ ($\downarrow \uparrow$, F ثابت)

$\downarrow P \uparrow v \Rightarrow \uparrow v \downarrow \tau \quad \downarrow X$ $\boxed{\downarrow \tau = \frac{V}{v \uparrow}}$ $\Rightarrow$ the conversion will be decrease

(e) Plot the rate of reaction and conversion as a function of PFR volume



ⓐ case ①



* for CSTR:

$X = 0.3 \Rightarrow \frac{F_A}{-r_A} = 800 \Rightarrow 0.3 * 800 = 240\ L$

$0.4 * \cancel{1000}\ 1200 = 480\ L$

Interpolation

$X = 0.366$

$\boxed{X_1 = 0.37}$

* for PFR:

$X = 0.45 \Rightarrow \frac{F_A}{-r_A} = 1250 \Rightarrow \underbrace{(0.45 - 0.4)(1250)}_{60} + \frac{1}{2}(0.45 - 0.4) * (1400 - 1200)$

مساحة مستطيل   مساحة مثلث

$60 + 5 = \textcircled{65}\ L$

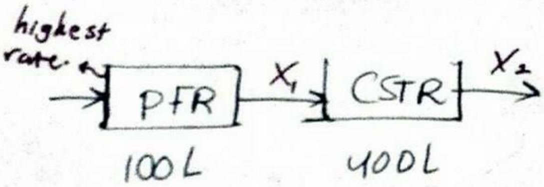
$X = 0.50 \Rightarrow \frac{F_A}{-r_A} = 1400 \Rightarrow (0.5 - 0.4)(1250) + \frac{1}{2}(0.5 - 0.4)(1600 - 1200)$

$120 + 15 = \textcircled{135\ L}$

Interpolation

$\boxed{X_2 = 0.48}$

case ②



* for PFR:

$X_1 = 0.2 \Rightarrow \frac{F_A}{-r_A} = 600 \Rightarrow \frac{1}{2} * 0.2 * 600 = 60\ L$

$= 0.3 \Rightarrow \frac{F_A}{-r_A} = 580 \Rightarrow \frac{1}{2} * 0.3 * 900 = 135\ L$

$\boxed{X_1 = 0.25}$

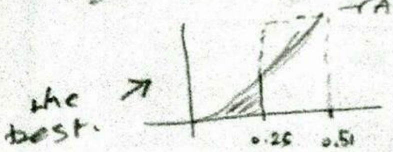
* for CSTR:

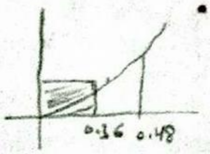
$X_2 = 0.5 \Rightarrow \frac{F_A}{-r_A} = 1500 \Rightarrow (0.5 - 0.25) * 1500 = 375\ L$

$X_2 = 0.6 \Rightarrow \frac{F_A}{-r_A} = 1800 \Rightarrow (0.6 - 0.25) * 1410 = 630\ L$

$X_2 = 0.509$

$X_2 = 0.51$





the case ② better than case ①

ⓒ $X = 0.5,\ \frac{F_A}{-r_A} = 750 \Rightarrow 0.5 * 750 = 375\ L$

$X = 0.55,\ \frac{F_A}{-r_A} = 790 \Rightarrow 0.55 * 790 = 434.5$

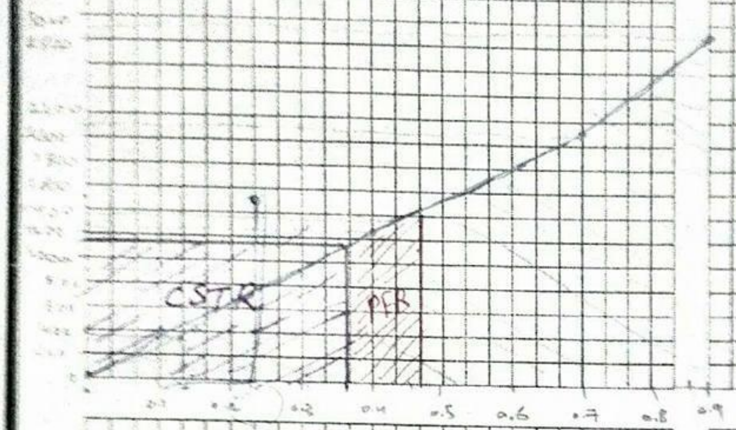
$\boxed{X = 0.52}$

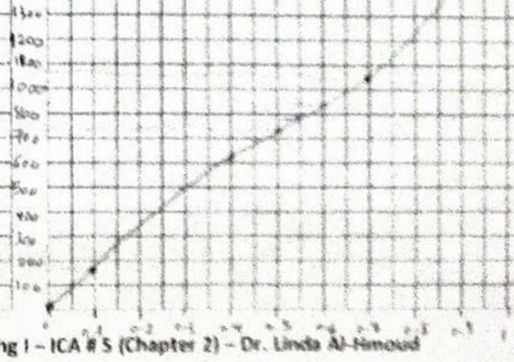
at $X_1 = 0.2 \rightarrow \frac{F_{A0}}{-r_A} = 480 \rightarrow V = 0.2 * 480 = 96$ L

$X = 0.4 \rightarrow$ $V = 500$ L

$X_1 = 0.37 \Rightarrow V = 360$ L

$X_2 = 0.42 \Rightarrow V = 100$ V





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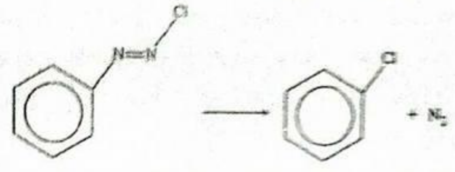
(0905421) Chemical Reaction Engineering I	**In-Class Assessment # 10 (Chapter 3)**
Partner (1) Name: رغد تيسير دويك	**Partner (2) Name:** علا الكباكي
Identity (1) #: 0153646	**Identity (2) #:**

Using the data in the table below, calculate the activation energy for the first order decomposition of benzene diazonium chloride to give chlorobenzene and nitrogen:

$k \rightarrow$ Arh
fn (Temp)
independent

اكثر من نقطتين ≠ [illegible]



k (s^{-1})	0.00043	0.00103	0.00180	0.00355	0.00717
T (K)	313.0	319.0	323.0	328.0	333.0
ln k	−7.75	−6.88	−6.32	−5.64	−4.94
1/T	3.2×10^{-3}	3.14×10^{-3}	3.1×10^{-3}	3.05×10^{-3}	3×10^{-3}

$\times 10^{-3}$ (1/T)	ln(K)
3.1	−6.5
3.15	−7.20

$\times 10^{-3}$ (1/T)	ln(k)
3.16	−7.20
3.06	−5.8

$\boxed{\text{slope} = -\frac{E}{R}}$

slope = −14

ما حنبدأ من (0,0)
لانه مو طالب ال intersept $\underline{(\ln A)}$
طالب ال {activation energy} from slope

$$\text{slope} = \frac{-7.20 + 6.5}{3.15 - 3.1}$$

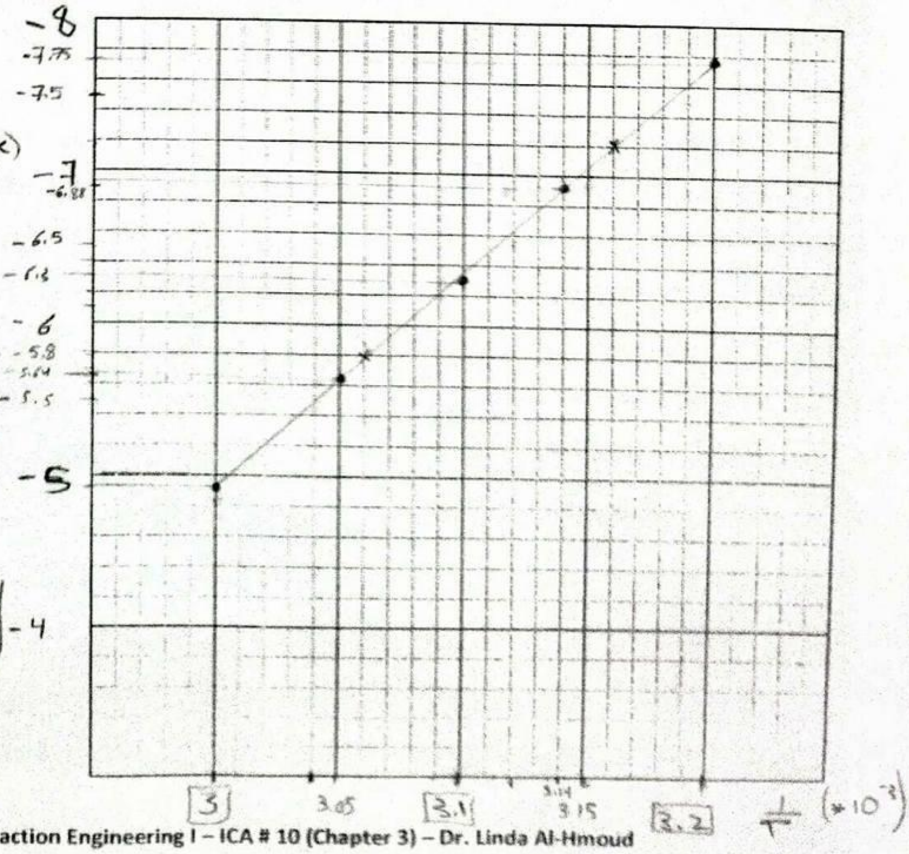
$$\text{slope} = -14$$

$$-14 = -\frac{E}{R}$$

$$-14 = \frac{-E}{8.314}$$

$$E = 14 \times 8.314$$

$$\boxed{E = 116.396 \text{ kJ/mol}}$$



Faculty of Engineering & Technology
Chemical Engineering Department

(0905421) Chemical Reaction Engineering I **First Semester – 2016/2017**

Pop Quiz # 1 (Chapter 3)

Name: [illegible] **ID #** 0152046

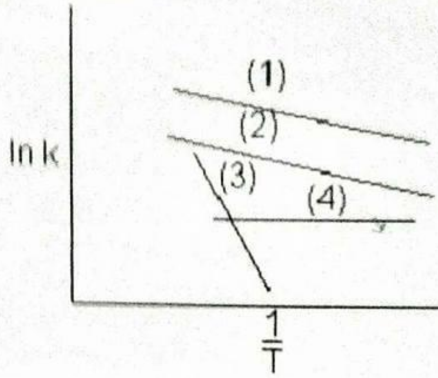
Consider the following elementary reactions

1) $A \xrightarrow{k_1} B$

2) $A \xrightarrow{k_2} D$

3) $A \xrightarrow{k_3} Y$

4) $A \xrightarrow{k_4} U$



$\ln k_A = \ln A - \frac{E}{R}\left(\frac{1}{T}\right)$

↑ slope ↑E

slop=0 K [illegible] independent on temp.

1) Which reaction is virtually temperature insensitive? (1) (2) (3) [✓ (4)]

2) Which reactions have the same activation energy? → Same Slope 1 AND 2

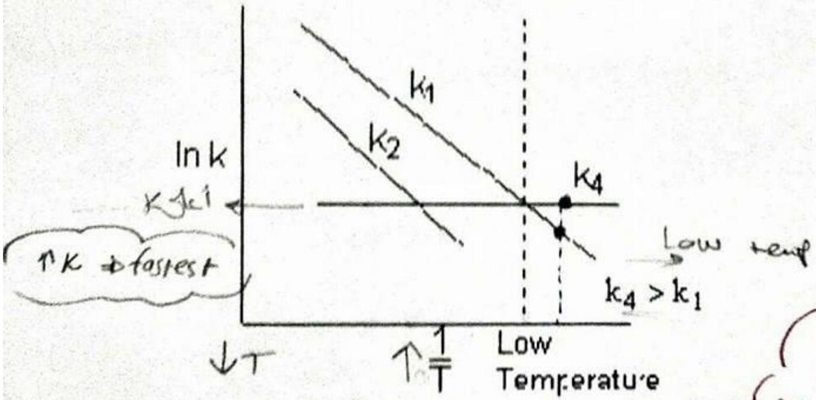
3) Which reaction will dominate (i.e. take place the fastest) at

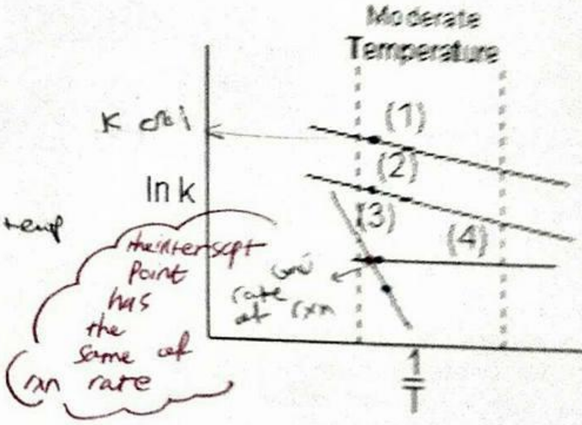
(a) low temperatures?

(1) (2) (3) [✓ (4)]

(b) moderate temperatures?

[✓ (1)] (2) (3) (4)



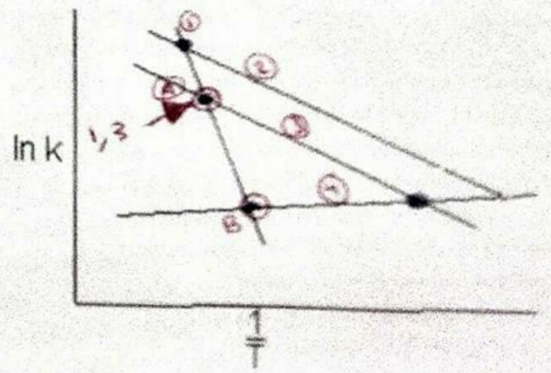


4) Label 2 intersecting points and complete the following sentences:

[illegible] rate of rxn

1- At point A, reaction 1 and reaction 3 will have the same rate of reaction at a given temperature.

2- At point B, reaction 1 and reaction 4 will have the same rate of reaction at a given temperature.



⇒ assume NH_3 is the limiting reactant

③ $\boxed{NH_3 = A}$

$$NH_3 + \frac{5}{4}O_2 \longrightarrow NO + \frac{3}{2}H_2O$$

Species	Symbol	Initial	Change	Remaining
NH_3	A	N_{A0}	$-N_{A0}X$	$N_{A0}(1-X)$
O_2	B	$N_{B0} = N_{A0}\Theta_B$	$-\frac{5}{4}N_{A0}X$	$N_{A0}(\Theta_B - \frac{5}{4}X)$
NO	C	—	$N_{A0}X$	$N_{A0}X$
H_2O	D	—	$\frac{3}{2}N_{A0}X$	$\frac{3}{2}N_{A0}X$
N_2	I	$N_I = \Theta_I N_{A0}$	—	$\Theta_I N_{A0}$

↙ inert

بنحطها نسبه
بالجدول

$$\Theta_B = \frac{C_{B0}}{C_{A0}} = \frac{y_{B0}}{y_{A0}} = \frac{(0.21)(0.85)}{0.15} = 1.2$$

$$\Theta_I = \frac{(0.79)(0.85)}{0.15} = 4.5$$

ⓐ

$$C_A = \frac{N_A}{V_0} = C_{A0}(1-x)$$

$$C_B = C_{A0}(\Theta_B - \frac{5}{4}x)$$

$$C_C = C_{A0}X$$

$$C_I = C_{A0}\Theta_I$$

$$C_D = \frac{3}{2}C_{A0}x$$

$$C_T = C_{A0} + C_{A0}\Theta_B - \frac{5}{4}C_{A0}X + \frac{3}{2}C_{A0}X + \Theta_I C_{A0}$$

$$= C_{T0} + \underbrace{\tfrac{1}{4}}_{1+\frac{3}{2}-1-\frac{5}{4}=\frac{1}{4}} C_{A0}X$$

$$P_T = C_T R T_0$$

$$P_T = (C_{T0} + \frac{1}{4}C_{A0}x)RT_0 \quad ; C_{A0} = 0.03 \text{ mol/L}$$

$$P_T = (0.2 + \frac{1}{4}(0.03)x)(0.08206)(500)$$

$$P_T = (0.2 + 7.5\times10^{-3}x)(41.03)$$

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(0905421) Chemical Reaction Engineering I	In-Class Assessment # 9 (Chapter 3)
Partner (1) Name: رغدا محمد البدوي	Partner (2) Name: تالا [illegible]
Identity (1) #: ____	Identity (2) #: ____

Production of Nitric acid

Nitric acid is made commercially from nitric oxide. Nitric oxide is produced by the gas-phase oxidation of ammonia.

mol fraction

$$4NH_3 + 5O_2 \rightarrow 4NO + 6H_2O$$

The feed consists of 15 mol% ammonia in air at 8.2 atm and 227 °C.

air → N_2 0.79, O_2 0.21

1. What is the total entering concentration?
2. What is the entering concentration of ammonia?
3. Set up a stoichiometric table with ammonia as your basis of calculation. no ε
   a. Express C_i for all species as functions of conversion for a constant-volume batch reactor. → NA
      Express P_T as a function of X.
   b. Express P_i and C_i for all species as functions of conversion for a flow reactor. → gas-phase (non-const. volume)
4. Write the combined rate law solely in terms of the ~~molar flow rates~~ conversion and rate law parameters for 3(a) and 3(b) above. Assume the reaction is first order in both reactants, and the specific reaction rate at the given condition is 5000 $L/(mol.min)$. ← k_A

NH_3	O_2
$\frac{0.15}{4}$	$\frac{0.85 * 0.21}{5}$
$\boxed{0.0375}$	$\boxed{0.0357}$
~~NH_3 is the limiting reactant~~	O_2 is the limiting reactant

$$NH_3 + \frac{5}{4} O_2 \rightarrow NO + \frac{6}{4} H_2O$$

$P = CRT$

$8.2 \text{ atm} = C * 0.08206 * (227 + 273)$

$C = 0.1998$

$C \simeq 0.20$

① $C_{T0} = \frac{P_0}{RT_0} = 0.2 \text{ mol/L}$

② $C_{A0} = y_{A0} C_{T0} = 0.15 (0.2) = 0.03 \text{ mol/L}$

$C_{A0} = y_{A0} C_{T0}$, $\frac{P}{RT}$

⇒ assume NH_3 is the limiting reactant

③ $\boxed{NH_3 = A}$

$$NH_3 + \frac{5}{4}O_2 \longrightarrow NO + \frac{3}{2}H_2O$$

Species	Symbol	Initial	Change	Remaining
NH_3	A	N_{A0}	$-N_{A0}X$	$N_{A0}(1-X)$
O_2	B	$N_{B0} = N_{A0}\Theta_B$	$-\frac{5}{4}N_{A0}X$	$N_{A0}(\Theta_B - \frac{5}{4}X)$
NO	C	—	$N_{A0}X$	$N_{A0}X$
H_2O	D	—	$\frac{3}{2}N_{A0}X$	$\frac{3}{2}N_{A0}X$
N_2 ↙ inert بنحطا شبه بالجدول	I	$N_I = \Theta_I N_{A0}$	—	$\Theta_I N_{A0}$

$$\Theta_B = \frac{C_{B0}}{C_{A0}} = \frac{y_{B0}}{y_{A0}} = \frac{(0.21)(0.85)}{0.15} = 1.2$$

$$\Theta_I = \frac{(0.79)(0.85)}{0.15} = 4.5$$

ⓐ

$$C_A = \frac{N_A}{V_0} = C_{A0}(1-X)$$

$$C_B = C_{A0}(\Theta_B - \frac{5}{4}X)$$

$$C_C = C_{A0}X$$

$$C_I = C_{A0}\Theta_I$$

$$C_D = \frac{3}{2}C_{A0}X$$

$$C_T = C_{A0} + C_{A0}\Theta_B - \frac{5}{4}C_{A0}X + \frac{3}{2}C_{A0}X + \Theta_I C_{A0}$$

$$= C_{T0} + \boxed{\frac{1}{4}}C_{A0}X \qquad \rightarrow 1 + \frac{3}{2} - 1 - \frac{5}{4} = \frac{1}{4}$$

$$P_T = C_T R T_0$$

$$P_T = (C_{T0} + \frac{1}{4}C_{A0}X)RT_0 \quad ; C_{A0} = 0.03 \text{ mol/L}$$

$$P_T = (0.2 + \frac{1}{4}(0.03)X)(0.08206)(500)$$

$$P_T = (0.2 + 7.5\times10^{-3}X)(41.03)$$

$$A + \frac{5}{4}B \rightarrow C + \frac{3}{2}D$$

3.b

$$C_A = \frac{C_{A0}(1-x)}{(1+\varepsilon x)} \underset{① \text{ isobaric}}{\cancel{\frac{P}{P_0}}} \underset{① \text{ isothermal}}{\cancel{\frac{T_0}{T}}}$$

$$\varepsilon = y_{A0}\,\delta$$

$$\delta = 1+\frac{3}{2} - 1 - \frac{5}{4} = \frac{1}{4}$$

$$\varepsilon = (0.15)\left(\frac{1}{4}\right) = 0.0375$$

$$C_{A0} = 0.03 \text{ mol/L}$$

$$P_A = C_A RT \qquad T = 500\text{ K}$$

$$C_B = \frac{C_{A0}(\Theta_B - \frac{5}{4}x)}{(1+\varepsilon x)} = \frac{0.03(1.2 - \frac{5}{4}x)}{(1+0.0375)}$$

$$P_B = C_B RT$$

$$C_C = \frac{C_{A0}X}{(1+\varepsilon x)}$$

$$P_C = C_C RT$$

$$C_D = \frac{\frac{3}{2}C_{A0}X}{(1+\varepsilon x)}$$

$$P_D = C_D RT$$

$$C_I = \frac{C_{A0}\Theta_I}{(1+\varepsilon x)}$$

$$P_I = C_I RT$$

④ $-r_A = k\,\underline{C_A}\,\underline{C_B}$

3(a) $\Rightarrow$ $-r_A = k C_{A_0}(1-X)\,C_{A_0}\left(\Theta_B - \frac{5}{4}X\right)$ $\Rightarrow$ const. Volume

3(b) $\Rightarrow$ $-r_A = k C_{A_0}^{2}\,\dfrac{(1-X)\left(\Theta_B - \frac{5}{4}X\right)}{(1+\varepsilon X)^{*}}$ $\Rightarrow$ not const. volume.

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(0905421) Chemical Reaction Engineering I	**In-Class Assessment # 11 (Chapter 3)**
Partner (1) Name: رغد الكسبري	**Partner (2) Name:** دانا كيالي
Identity (1) #:	**Identity (2) #:**

P3-11$_A$ Set up a stoichiometric table for the each of the following reactions and express the concentration of each species in the reaction as a function of conversion evaluating all constants (e.g. ε. Θ). Then, assume the reaction follows an elementary rate law, and write the reaction rate solely as a function of conversion, i.e., $-r_A = f(X)$.

(c) For the isothermal, isobaric, catalytic gas-phase oxidation

$\downarrow$ Non-constant volume

→NO limiting reactant

the feed enters a PBR at 6 atm and 260°C and is a stoichiometric mixture of only oxygen and ethylene.

↳ $\Theta_B = \frac{1/2}{1} = 1/2$

↳ $y_{A0} = \frac{f_{B0}}{f_{A0}+f_{B0}} = \frac{1}{(1+\frac{1}{2})}$

Just from stoichiometric

عشان ما عنا ، مقدام
دخلت بالزبط $\frac{1}{2}O_2$

$$\underbrace{C_2H_4 + \frac{1}{2}O_2}_{\text{limiting}} \rightarrow CH_2OCH_2$$

① Stoich. table

Species	Symbol	fed	change	effluent
C_2H_4	A	F_{A0}	$-F_{A0}X$	$F_{A0}(1-X)$
O_2	B	$F_{B0}=F_{A0}\Theta_B$	$-\frac{1}{2}F_{A0}X$	$F_{A0}(\underset{\frac{1}{2}}{\Theta_B}-\frac{1}{2}X)$
CH_2OCH_2	C	$F_{C0}=0$	$+F_{A0}X$	$F_{A0}X$

$\boxed{\Theta_B = \frac{1}{2}}$

$$\ast\ C_A = \frac{F_A}{\nu} = \frac{F_{A0}(1-X)}{\nu_0(1+\varepsilon X)} = C_{A0}\frac{(1-X)}{(1-\frac{X}{3})}$$

$$\ast\ C_B = \frac{C_{A0}}{2}\frac{(1-X)}{(1-\frac{X}{3})}$$

$$C_C = \frac{C_{A0}X}{(1-X/3)}$$

$$\varepsilon = \delta\, \overset{\text{from stoich}}{y_{A0}}$$

$$= \left(1-\frac{1}{2}-1\right)\left(\frac{1}{\underset{Ⓐ}{1}+\underset{Ⓑ}{\frac{1}{2}}}\right)$$

$$= -\frac{1}{3}$$

$$-r_A = k\,C_A\,C_B^{1/2}$$

$$-r_A = k\,\frac{C_{A0}(1-X)}{1-\frac{X}{3}}\left(\frac{\frac{C_{A0}}{2}(1-X)}{1-\frac{X}{3}}\right)^{1/2}$$

$$C_{A0} = \frac{y_{A0}P_0}{RT} = \frac{\frac{2}{3}\times 6\text{ atm}}{0.08206\times(260+273)}$$

$$= 0.091\text{ mol/dm}^3$$

P3-16$_B$ Calculate the equilibrium conversion and concentrations for each of the following reactions.

X_e

(a) The liquid-phase reaction

↳ const. volume

$$A + B \leftrightarrow C \quad \Leftarrow \text{reversible rxn.}$$

with $C_{A0} = C_{B0} = 2$ mol/dm^3 and $K_C = 10$ dm^3/mol.

$$C_{A0} = \frac{N_0}{V_0}$$

$$-r_A = k\left(C_A C_B - \frac{C_C}{k_c}\right) \overset{\text{at equilibrium}}{=} 0 \Rightarrow C_A{}^2(1-x_e)^2 - \frac{C_{A0} x_e}{K_c} = 0$$

$$\Rightarrow \boxed{C_A = C_{A0}(1-x)} \qquad (1-x_e)^2 - \frac{x}{2(10)} = 0$$

$$C_B = C_{A0}(\Theta_B - x), \quad \Theta_B = 1 \qquad \Rightarrow \boxed{X_e = 0.8}$$

$$\Rightarrow \boxed{C_B = C_{A0}(1-x)}$$

$$C_{Ae} = 2(1-0.8) = 0.4 \text{ mol/L} = C_{Be}$$

$$C_C = C_{A0}(\Theta_C + x)$$

$$\Rightarrow \boxed{C_C = C_{A0} X} \qquad C_{Ce} = 2(0.8) = 1.6 \text{ mol/L}$$

(b) The gas-phase reaction

↳ non-const. volume

$$A \leftrightarrow 3C \quad \text{↳ reversible}$$

carried out in a flow reactor with no pressure drop. Pure A enters at a temperature of 400 K and 10 atm. At this temperature, $K_C = 0.25(\text{dm}^3/\text{mol})^2$.

$$-r_A = k\left(C_A - \frac{C_C^3}{k_c}\right)$$

$$C_{A0} = \frac{y_{A0} P_0}{R T_0} = \frac{(1)(10 \text{ atm})}{0.08206 \frac{\text{atm}\cdot\text{L}}{\text{mol}\cdot\text{K}} \times 400 \text{ K}} = 0.305 \text{ mol/L}$$

$$C_A = \frac{C_{A0}(1-x)}{(1+\varepsilon x)} =$$

Pure A

$$\varepsilon = y_{A0}\,\delta = (2)(1) = 2$$

$$C_C = \frac{C_{A0}(\Theta_C + 3x)}{(1+\varepsilon x)}, \quad \Theta_C = 0$$

Pure A enters $\boxed{F_{C0} = 0}$

$$= \frac{3 C_{A0} X}{1 + \varepsilon X}$$

at equilibrium $-r_A = 0$

$$\frac{C_{A0}(1-X_e)}{1+2X_e} - \frac{(3\,C_{A0} X_e)^3}{k_c(1+2X_e)^3} = 0 \Rightarrow X_e = 0.58$$

$$C_{Ae} = \frac{(0.305)(1-0.58)}{1+2(0.58)} = 0.06 \text{ mol/L}$$

$$C_{Ce} = \frac{3(0.305)(0.58)}{1+2(0.58)} = 0.25 \text{ mol/L}$$

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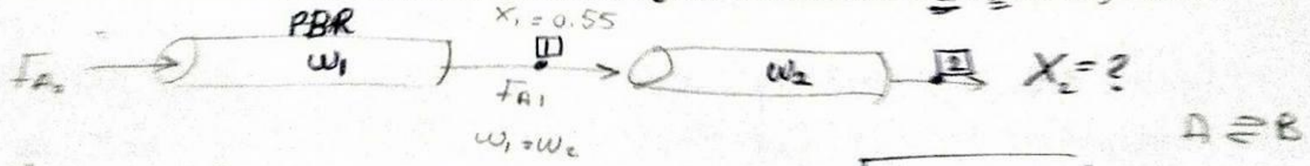
(0905421) Chemical Reaction Engineering I	**In-Class Assessment # 13 (Chapter 4)**
Partner (1) Name: رغد البدوي	**Partner (2) Name:**
Identity (1) #:	**Identity (2) #:**

P4-13 Compound A undergoes a reversible isomerization reaction, $A \rightleftarrows B$, over a supported metal catalyst. Under pertinent conditions, A and B are liquid, miscible, and of nearly identical density; the equilibrium constant for the reaction (in concentration units) is 5.8. In a *fixed-bed isothermal flow reactor* in which back mixing is negligible (i.e. plug flow), a feed of pure A undergoes a net conversion to B of 55%. The reaction is elementary. If a second, identical flow reactor at the same temperature is placed downstream from the first, what overall conversion of A would you expect if:

(a) The reactors are directly connected in series? *(Ans.: X = 0.745.)*
(b) The products from the first reactor are separated by appropriate processing and only the unconverted A is fed to the second reactor?

Solution (a): *Follow the algorithm step-by-step:*

0- Draw a sketch of the problem and write down all given information: *K_C, X_1 and any others.*



$$X_2 = \frac{\text{total moles reacted at Point 2}}{\text{mole fed to first reactor}}$$

$$\boxed{K_c = 5.8}$$

1- Write the **mole balance** equation. Remember, the mole balance depends on the ***reactor type***.

$$\frac{dx}{dw} = \frac{-r_A'}{F_{A0}}$$

2- Write the **rate law**. Ask yourself: is it elementary or does it follow elementary rate law?

$$-r_A = k\left[C_A - \frac{C_B}{K_c}\right] \Leftarrow \text{elementry}$$

3- Do the **stoichiometry (i.e. write the concentration of species appearing in the rate law in terms of conversion)**. Ask yourself again: Is it [✓ *liquid*] or *gas phase*? [✓ ***Constant volume***] or not?

Lq. $v = v_0$ → No ε

$$C_A = \frac{F_A}{v_0} = \frac{F_{A0}(1-x)}{v_0} = \underline{C_{A0}(1-x)}$$

$$C_B = \frac{F_B}{v_0} = \frac{F_{A0}x}{v_0} = \underline{C_{A0}x} \quad ; \quad \underset{\text{Product}}{\Theta_B} = \frac{F_{B0}}{F_{A0}} = \boxed{0}$$

4- Now the last step: **COMBINE**. This means putting the concentrations in step (3) back in the rate law (2), and substituting this rate law, which is now in terms of conversion, in the mole balance you wrote in (1).

$$-r_A = kC_{A_0}\left[1 - X - \frac{X}{K_c}\right]$$

$$\int_0^{\omega_1} d\omega = \int_0^{X_1} \text{---}$$

$$\int_0^{\omega_2} \partial\omega = \int_{X_1}^{X_2} \text{---}$$

$$\checkmark\ \frac{dX}{d\omega} = \frac{-r_A'}{F_{A_0}} = \frac{kC_{A_0}}{F_{A_0}}\left[1 - X\left(1 + \frac{1}{K_c}\right)\right]$$

5- Finally, solve the equation you got in (4) and find X_2. $\Rightarrow \omega_1 = \omega_2$

$$\boxed{\underline{\underline{\omega_1}} = \frac{F_{A_0}}{kC_{A_0}}\int_0^{X_1}\frac{dX}{\left[1 - \left(1 + \frac{1}{K_c}\right)X\right]}} = \boxed{\underline{\underline{\omega_2}} = \frac{F_{A_0}}{kC_{A_0}}\int_{X_1}^{X_2}\frac{dX}{\left[1 - \left(1 + \frac{1}{K_c}\right)X\right]}}$$

$$\omega_1 = \frac{F_{A_0}}{\cancel{k}C_{A_0}}\,\frac{1}{\cancel{\left(1 + \frac{1}{K_c}\right)}}\ln\left(\frac{1}{1 - \left(1 + \frac{1}{K_c}\right)X_1}\right)$$

$$\boxed{\div}$$

$$\omega_2 = \frac{\cancel{F_{A_0}}}{kC_{A_0}}\,\frac{1}{\cancel{\left(1 + \frac{1}{K_c}\right)}}\ln\frac{\left(1 - \left(1 + \frac{1}{K_c}\right)X_1\right)}{\left(1 - \left(1 + \frac{1}{K_c}\right)X_2\right)}$$

cancel unknowns F_{A_0}, k and C_{A_0}

$$1 = \frac{\ln\left(\frac{1}{(1 - 1.172X_1)}\right)}{\ln\left(\frac{1 - 1.172X_1}{1 + 1.172X_2}\right)}$$

$$X_2 = \frac{\left[1 - \left(1 - \left(1 + \frac{1}{K_c}\right)X_1\right)^2\right]}{1 + \frac{1}{K_c}}$$

$$e^{\ln\left(\frac{1 - 1.172X_1}{1 + 1.172X_2}\right)} = e^{\ln\left(\frac{1}{1 - 1.172X_1}\right)}$$

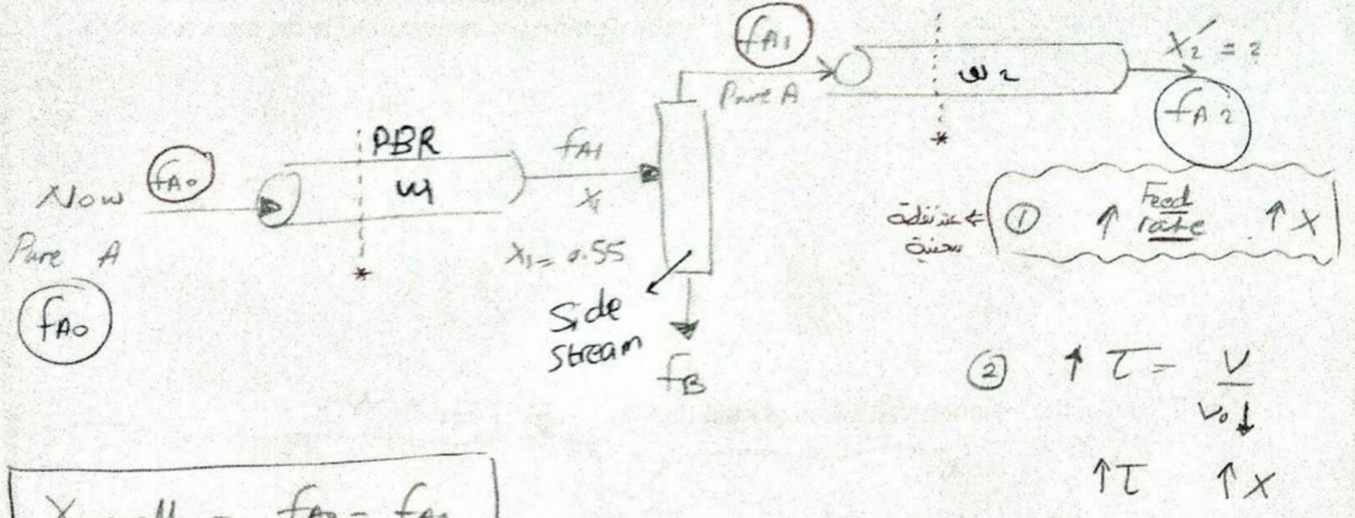
Substitute $X_1 = 0.55$ and $K_c = 5.8$

$$-1\,(1 - 1.172X_1)^2 = 1 + 1.172X_2\ {-1}$$

$$\boxed{X_2 = 0.745}$$

$$\boxed{\frac{1 - (1 - 1.172X_1)^2}{1.172} = X_2}$$

Part ②



$$\boxed{X_{overall} = \frac{f_{A0} - f_{A2}}{f_{A0}}}$$

⇒ conversion (X_2') based on $\underline{f_{A1}}$

$$W_1 = \frac{\cancel{f_{A0}}}{\cancel{kC_{A0}}} \frac{1}{\cancel{(1+\frac{1}{K_c})}} \ln\left(\frac{1}{1-(1+\frac{1}{K_c})X_1}\right)$$

÷

$$W_2 = \frac{f_{A1}}{kC_{A0}} \int_0^{X_2'} \frac{dx}{(1-(1+\frac{1}{K_c})x)} = \frac{\cancel{f_{A0}}(1-X_1)}{\cancel{kC_{A0}}\cancel{(1+\frac{1}{K_c})}} \ln\left(\frac{1}{(1-(1+\frac{1}{K_c})X_2')}\right)$$

$$W_1 = W_2$$

$$\boxed{f_{A1} = f_{A0}(1-X_1)}$$

$$1 = \frac{1}{(1-X_1)} \frac{\ln\left(\frac{1}{1-1.172X_1}\right)}{\ln\left(\frac{1}{1-1.172X_2'}\right)}$$

⇒ Substitute for fA1 and cancel fA0, CA0, k

One equation and one unknown solving for X_2'

$$\boxed{X_2' = 0.768}$$

$$e^{\ln\left(\frac{1}{1-1.172X_2'}\right)} = e^{\left[\frac{1}{(1-X_1)} \ln\left(\frac{1}{1-1.17X_1}\right)\right]}$$

$$\frac{1}{1-1.172X_1} = e^{\frac{1}{(1-X_1)}} + \frac{1}{1-1.172X_2'}$$

⇒ $\underline{X_{overall}} = \frac{f_{A0} - f_{A2}}{f_{A0}} = \frac{f_{A0} - f_{A1}(1-X_2')}{f_{A0}}$

$$= \frac{\cancel{f_{A0}} - \cancel{f_{A0}}(1-X_1)(1-X_2')}{\cancel{f_{A0}}}$$

$$\boxed{0.896} = 1 - (1-0.55)(1-0.77) \leftarrow 1 - (1-X_1)(1-X_2')$$

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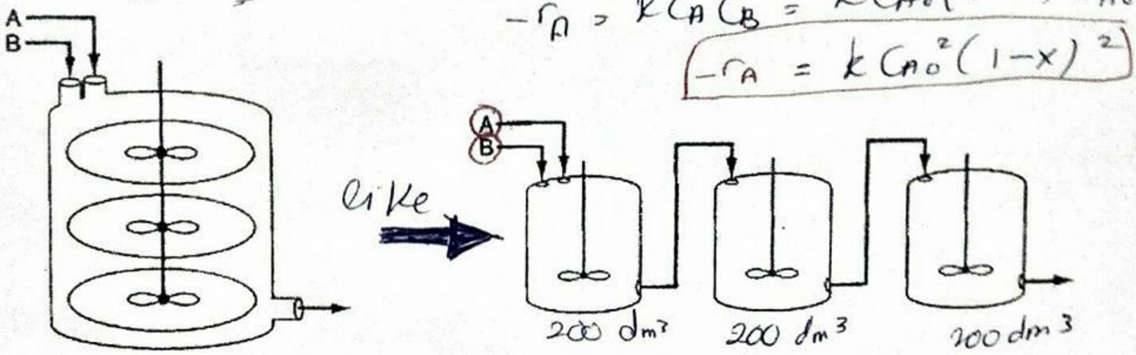
(0905421) Chemical Reaction Engineering I	**In-Class Assessment # 15 (Chapter 4)**
Partner (1) Name: [illegible]	**Partner (2) Name:** ________
Identity (1) #: ________	**Identity (2) #:** ________

→ const. volume

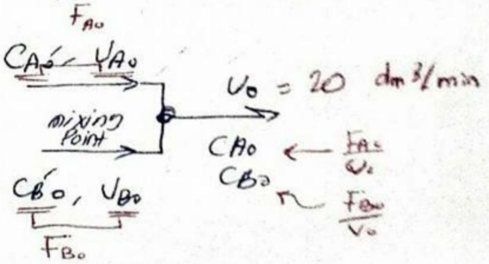
The elementary liquid-phase reaction $A + B \rightarrow C$ is to be carried out in a CSTR with three impellers. The mixing patterns in the CSTR are such that it is modeled as three equal-sized CSTRs in series. Species A and B are fed in separate lines to the CSTR, which is initially filled with inert material. Each CSTR is 200 dm³ and the volumetric flow to the first reactor is 10 dm³/min of A and 10 dm³/min of B.

$-r_A = k C_A C_B = k C_{A0}(1-x) C_{A0}(1-x)$

$\boxed{-r_A = k C_{A0}^2 (1-x)^2}$



A. What is the conversion of A?
B. If the three tanks were connected in parallel with A and B split equally among the three reactors, what will be the conversion of A?



Additional Information
$C_{A0}' = C_{B0}' = 2.0$ mol/dm³
$k = 0.025$ dm³/mol/min

$v_{A0} = 10$, $v_{B0} = 10$

$v_0 = 10 + 10 = 20$ dm³/min

$$C_{A0} = \frac{F_{A0}}{v_0} = \frac{(C_{A0}')(v_{A0})}{v_0}$$

$v_0 = v_{A0} + v_{B0} = 20$ dm³/min

$$C_{A0} = \frac{F_{A0}}{v_0} = \frac{\overset{C_{A0}'}{②} \times \overset{v_{A0}}{10} \text{ mol/min}}{20 \text{ dm}^3/\text{min}} = \underline{1 \text{ mol/dm}^3} = C_{B0}$$

conversion for (n) reactors in series:

$$\boxed{X = 1 - \frac{1}{(1+D_a)^n}}$$

2nd order respect to Ⓐ

$$D_A = \overset{ⓚ}{k} \, \overset{(C_{A0})}{C_{A0}} \, \tau = 0.025 \frac{\text{dm}^3}{\text{mol.min}} \times \frac{1 \text{mol}}{\text{dm}^3} \times \frac{200}{20}$$

mol

V_0 total

$\boxed{D_A = 0.25}$ ✓

$$= 1 - \left(\frac{1}{(1+\underset{0.25}{[illegible]})^3}\right) = ~~0.4[illegible]~~ \boxed{0.488 = X}$$ ✓

⇒ only if they have same volume.

ⓑ $V_{total} = \frac{F_{A_0} X}{k\, C_{A_0}^2 (1-X)^2} =$

total
$200 \times 3 \rightarrow 600 = \frac{20X}{0.025(1)(1-X)^2} \equiv 200 = \frac{(20/3)\, X}{(0.025)(1)(1-X)^2}$

$\boxed{X = 0.333}$
Parallel

School of Engineering
Chemical Engineering Department

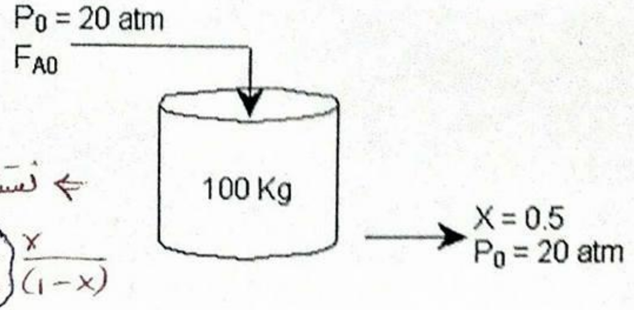
(0905421) Chemical Reaction Engineering I **In-Class Assessment # 16 (Chapter 4)**

Partner (1) Name: [illegible] **Partner (2) Name:** [illegible]

Partner (1) ID # : ______ **Partner (2) ID # :** ______

The elementary isomerization reaction $A \rightarrow B$ is carried out isothermally at 20 atm in a fluidized CSTR containing 100 kg of catalyst, where 50% conversion is achieved.

W X



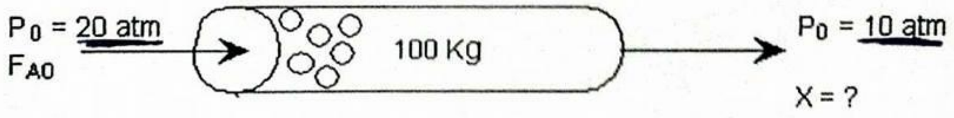
fluidized نستطيع أن is CSTR ⇐

$$\hookrightarrow W = \boxed{\frac{F_{A0}}{kC_{A0}}} \frac{X}{(1-X)}$$

It is proposed to replace the CSTR with a packed bed reactor

تم استبدال ال CSTR

PBR ب



The entering pressure was 20 atm and the exit pressure was found to be 10 atm.

A. What would be the conversion in the PBR if there were NO pressure drop?
Hint 1: What information can you obtain from the CSTR?

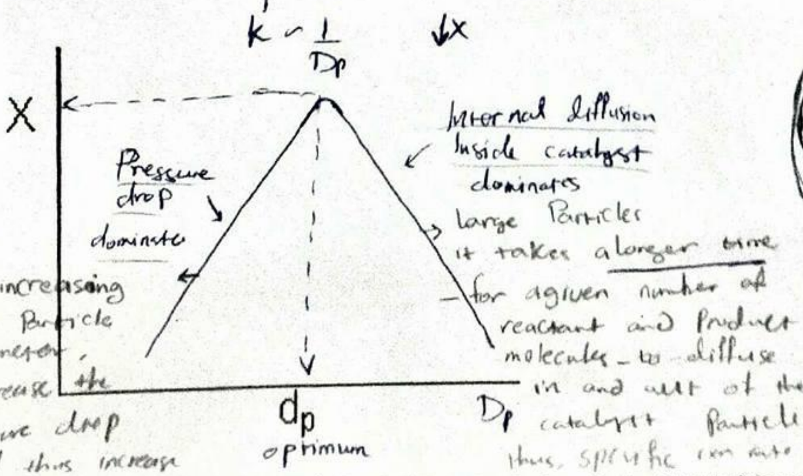
B. What will be the conversion in the new PBR with pressure drop?
Hint 2: How do you calculate the pressure drop parameter, α?

C. Plot the conversion as a function of particle diameter assume everything else is unchanged. Why does the X-d_P curve look the way it does?

(dp)

mass transfer (rate band

$\uparrow$ $k \sim \frac{1}{D_p}$ $\downarrow X$



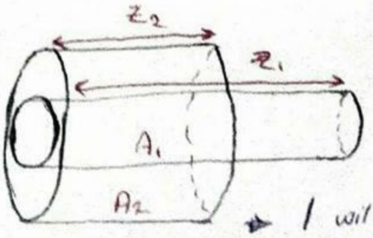
large Particles it takes a longer time for a given number of reactant and Product molecules to diffuse in and out of the catalyst Particle thus, specific rxn rate decrease with ↑Dp

By increasing the Particle diameter, decrease the Pressure drop and thus increase the rxn rate and the conversion

↑Dp ↓ΔP ↑-rA ↑X

↑Dp ↑time $k \sim \frac{1}{D_p}$ ↓k ↓X

D. Suppose you could choose another pipe diameter to hold the same amount of catalyst. Would you choose a larger or smaller pipe diameter, or would you use the current pipe diameter? Explain.



$A_1 < A_2$

$D_1 < D_2$

$W_1 = W_2$

→ I will choose the ~~[illegible]~~ Larger diameter; because the Pressure drop increase as distance (Z) increase. and decrease conversion

Classwork (16)

①

$A \rightarrow B$ elementery

$W = 100$ kg of catalyst

$P = 10$ atm
fluidized CSTR $\Rightarrow W = \frac{F_{A0}X}{-r'_A}$

① mole balance :

$$F_{A0}\frac{dx}{dw} = -r'_A$$

② rate law :

$$-r'_A = k\,C_A$$

③ Stoichiometry :

$$C_A = C_{A0}\frac{(1-X)}{(1+\varepsilon X)}\underset{\text{NO }\Delta P}{\cancel{y}}\ \underset{\text{isothermal}}{\cancel{\frac{T_0}{T_1}}} \quad ; \varepsilon = (1-1)y_{A0}$$

$$\boxed{\varepsilon = \text{zero}}$$

$$C_A = C_{A0}(1-X)$$

④ combine

$$-r'_A = k\,C_{A0}(1-X)$$

$$F_{A0}\frac{dx}{dw} = \underset{?}{\underline{\underline{k}}}\ \underset{?}{\underline{\underline{C_{A0}}}}(1-X)$$

$\Rightarrow$ From fluidized bed :

$$W = \frac{F_{A0}X}{-r'_A}$$

$$100 = \frac{F_{A0}(0.5)}{k\,C_{A0}(1-x)} \Rightarrow 100 = \frac{F_{A0}}{kC_{A0}}\ \frac{\cancel{0.5}}{(1-\cancel{0.5})} \Rightarrow \frac{F_{A0}}{kC_{A0}} = 100$$

$$\int_0^X \frac{dx}{(1-x)} = \frac{k\,C_{A0}}{F_{A0}}\int_0^W dw$$

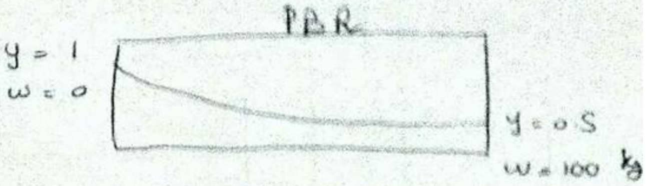
$$-\ln(1-x)\Big|_0^X = \frac{1}{100}\ W$$

$$-\ln(1-x) = \frac{100}{100}$$

$$-\ln(1-x) = 1$$

$$\boxed{X = 0.632}$$

(B) with Pressure drop (y)



① mole balance : $F_{A0}\frac{dx}{dw} = -r'_A$

② rate Law : $-r'_A = k\,C_A$

③ stoichiometry : $C_A = C_{A0}\frac{(1-x)}{(1+\epsilon x)}\,y\,\cancel{\frac{T_0}{T}}$ ϵ = zero

Isothermal

$$C_A = C_{A0}(1-x)\,y$$

④ combine : $-r'_A = k\,C_{A0}(1-x)\,y$

$$F_{A0}\frac{dx}{dw} = k\,C_{A0}(1-x)\,y \quad ; \quad y = \sqrt{1-\alpha w}$$

$$\int_0^x \frac{dx}{1-x} = \frac{k\,C_{A0}}{F_{A0}}\int_0^w y\,dw$$

$w = \frac{F_{A0}X}{-r'_A} = \frac{F_{A0}}{kC_{A0}}\frac{X}{(1-X)}$

$$-\ln(1-x)\Big|_0^x = \frac{1}{100}\int_0^w y\,dw$$

$100 = \frac{F_{A0}}{kC_{A0}}\frac{0.5}{(1-0.5)}$

$$-\ln(1-x) = \frac{1}{100}\int_0^w \frac{dw\,(1-\alpha w)^{1/2}}{\cancel{(1-\alpha w)^{1/2}}}$$

$100 = \frac{F_{A0}}{kC_{A0}}$

$\frac{F_{A0}}{kC_{A0}} = 100$

~~$-\ln(1-x) = \frac{1}{100}\,\frac{1}{(0.5)}\,\frac{(1-\alpha w)^{1/2}}{(-\alpha)}\Big|_0^w$~~

~~$-\ln(1-x) = \frac{1}{(50)(-\alpha)}\left((1-\alpha w)^{1/2}-1\right)$~~ ; $\alpha = ?$

$y = \sqrt{1-\alpha w}$

~~$-\ln(1-x) = \frac{1}{\underbrace{(50)(-7.5\times10^{-3})}_{-0.375}}\left(\left(1-\underbrace{(7.5*10^{-3})(100)}_{0.75}\right)^{1/2}-1\right)$~~

$\frac{10}{20} = \sqrt{1-\alpha(100)}$

$\boxed{\alpha = 7.5\times10^{-3}\ kg^{-1}}$

~~$-\ln(1-x) = +1.333$~~

on calcuater

$-\ln(1-x) = \frac{1}{100}(50)$ $\Rightarrow$ $\boxed{X = 0.393}$

Partner (1) Name: رغد البدوي **Partner (2) Name:** ____________

Partner (1) ID # : ____________ **Partner (2) ID # :** ____________

What four things are wrong with this membrane reactor solution?

$A \rightarrow \frac{1}{2}B + C$

$\varepsilon = \overset{\delta}{(\frac{1}{2}+\cancel{1}-\cancel{1})}\overset{y_{A0}}{(\frac{1}{2})} = \frac{1}{4}$

$\boxed{\varepsilon \neq 0}$

$\boxed{\varepsilon = 0.25}$

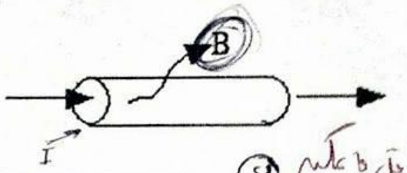
The gas phase elementary reaction

$(2A \rightarrow B + 2C)/2$

* ما بتكتب ال conversion لانه نغير بالحل في حالة ال (membrane) ولكن نكتب بدلالة ال molar flow rate (F)

$$C_{t0} = 0.4\,\frac{\text{mol}}{\text{dm}^3}$$

$$F_{A0} = \frac{10\,\text{mol}}{\text{s}}$$



sweep gas (inert gas), take Ⓑ ⇒ $C_{AS} = 0$

⑨ مستقيمة ، للتركيز على السطح أقل ما يمكن

is carried out in a membrane reactor in which there is pressure drop with $\alpha = 0.019\ \text{kg}^{-1}$. An equal molar feed of A and inerts, I, enter the reactor and only species B can exit the membrane. Plot the conversion of A down the membrane reactor containing 50 kg of catalyst. The specific reaction rate is $k_A = 2.5$ $\text{dm}^6/\text{mol/kg/s}$, and the mass transfer coefficient $k_C = 1.5\ \text{dm}^3/\text{kg}\cdot\text{s}$

Proposed Solution

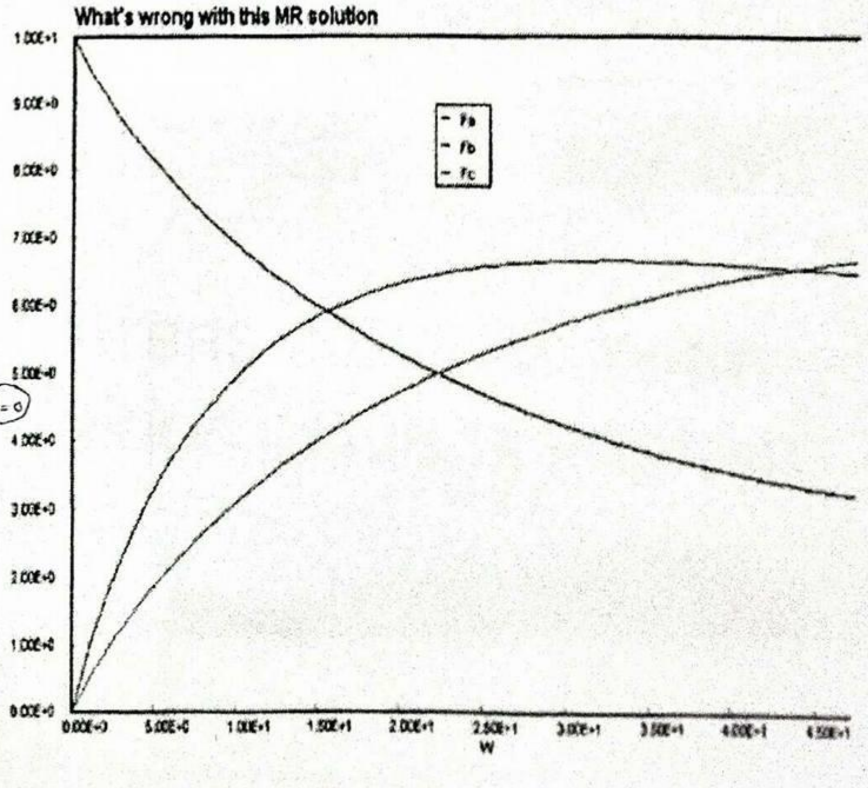
⇒ Differential equations

mole balance for each component.

1 d(Fa)/d(W) = ra ✓
2 d(Fb)/d(W) = rb-Rb ✓
3 d(Fc)/d(W) = rc ✓

⇒ Explicit equations

F_I × ① Ft = Fa+Fb+Fc +F_I
2 ka = 2.5 ✓
3 Cto = 0.4 ✓
4 alpha = 0.019 ✓
$\frac{dy}{dw} = \frac{-\alpha}{2y}\frac{F_T}{F_{T0}}$ × ⑤ y = (1-alpha*W)^0.5 ($\varepsilon = 0$)
6 Cb = Cto*Fb*y/Ft
7 Ca = Cto*Fa*y/Ft
$r'_B = \frac{r_A'}{-2} = \frac{kC_A^2}{2}$ ⑧ rb = ka*Ca
⑨ ra = -rb/2
10 rc = -ra ✓
11 kc = 1.5 ✓
12 Rb = kc*Cb ✓



① mole balance

$\boxed{\frac{dF_A}{dW} = r'_A}$ $\boxed{\frac{dF_B}{dW} = r'_B - R_B}$ $\boxed{\frac{dF_C}{dW} = r'_C}$

② rate law:

$-r_A = kC_A^2$ $R_B = k_c C_B$

$\boxed{r'_A = -k C_A^2}$

relative rate:

$$\boxed{\frac{r'_A}{-2} = \frac{r'_B}{1} = \frac{r'_C}{2}}$$

$$r'_B = \frac{r'_A}{-2} = \frac{r'_C}{2}$$

$$\boxed{R_B = k_c C_B}$$

③ Stoich.

$$\boxed{C_A = C_{T0} \frac{F_A}{F_T} y}$$

$$\boxed{C_B = C_{T0} \frac{F_B}{F_T} y}$$

$$\boxed{F_T = F_A + F_B + F_C + F_I}$$

$$\varepsilon = \underset{\downarrow}{y_A}\ \underset{\downarrow}{\delta}$$

$$= \left(\tfrac{1}{2}\right)\left(\tfrac{1}{2}\right)$$

$\varepsilon = 0.25 \neq$ zero

$$\frac{dy}{dW} = \frac{-\alpha}{2y} \frac{F_T}{F_{T0}}$$

⇒ evaluated Parameters

- ✓ $F_{A(0)} = 10$ mol/s
- ✓ $F_{B(0)} = 0$
- ✓ $F_{C(0)} = 0$
- ✓ $W(0) = 0$
- ✓ $W(f) = 50$ kg
- ✓ $y(0) = 1$
- ✓ $k = 2.5 \frac{dm^6}{mol \cdot kg \cdot s}$
- ✓ $k_c = 1.5 \frac{dm^3}{kg \cdot s}$
- ✓ $\alpha = 0.019\ kg^{-1}$
- ✓ $C_{T0} = 0.4$ mol/dm³
- ✓ $F_{I0} = 10$ mol/s

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(0905421) Chemical Reaction Engineering I	**In-Class Assessment # 18 (Chapter 5)**
Partner (1) Name: (غنا البدوي)	**Partner (2) Name:**
Partner (1) ID #: 0153646	**Partner (2) ID #:**

The reaction A → B is carried out in a constant volume batch reactor. Determine the reaction order and specific reaction rate from the following data.

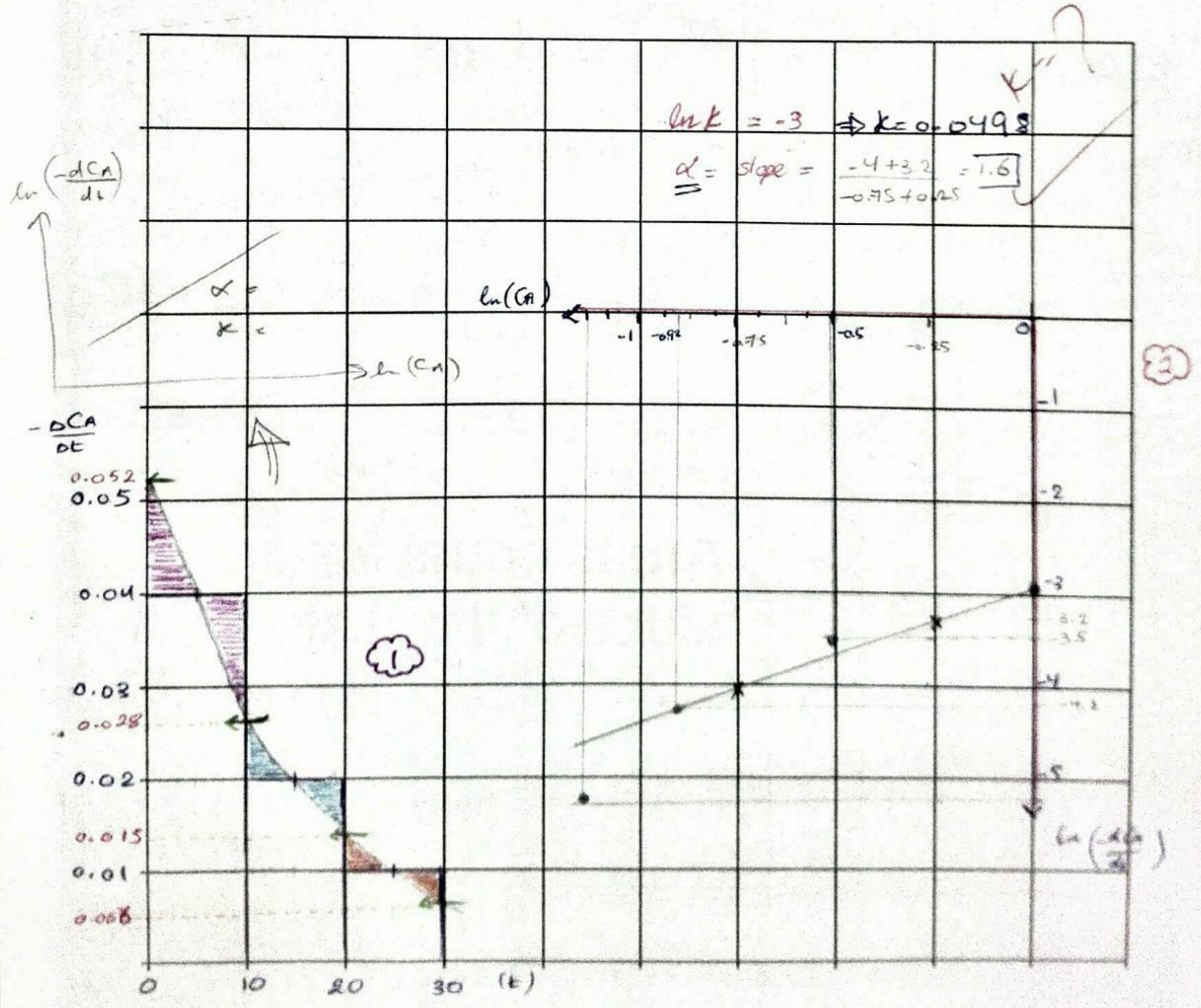
①

t (min)	0	10	20	30
C_A(mol/dm^3)	1	0.6	0.4	0.3
$-\Delta C_A/\Delta t$	0.04	0.02	0.01	
$-dC_A/dt$	0.052	0.028	0.015	0.008

$\ln(-dC_A/dt)$

②

t(min)	0	10	20	30
$\ln(C_A)$	0	-0.51	-0.92	-1.2
$-dC_A/dt$	0.05	0.03	0.015	0.00
$\ln(-dC_A/dt)$	-2.99	-3.5	-4.2	-5.3



$\ln k = -3 \Rightarrow k = 0.0498$

$\underline{\underline{\alpha}} = \text{slope} = \frac{-4+3.2}{-0.75+0.25} = \boxed{1.6}$

2. Numerical differentiation formulas

Time (min)	t_0	t_1	t_2	t_3	t_4	t_5
Concentration (mol/dm³)	C_{A0}	C_{A1}	C_{A2}	C_{A3}	C_{A4}	C_{A5}

ln C_A / ln (dC_A/dt)

The three-point differentiation formulas[1]

Initial point:

$$\left(\frac{dC_A}{dt}\right)_{t_0} = \frac{-3C_{A0} + 4C_{A1} - C_{A2}}{2\Delta t}$$

① → 0.6   0.4   0.3

$$\left(\frac{dC_A}{dt}\right)_{i=1} = \frac{-3(1) + 4(0.6) - 0.4}{2(10)} = -0.05$$

Interior points:

$$\left(\frac{dC_A}{dt}\right)_{t_{i=2}} = \frac{1}{2\Delta t}[(\underset{\substack{(2+1)=3\\ \text{third point}}}{C_{A(i+1)}} - \underset{\substack{(2-1)=1\\ \text{first point}}}{C_{A(i-1)})}]$$

2nd point

1 ← ⓪.6 → 0.4
0.6 ← (0.4) → 0.3
} Intermediate Points.

$$\left(\frac{dC_A}{dt}\right)_{i=2} = \frac{0.4 - 1}{2(10)} = -0.03$$

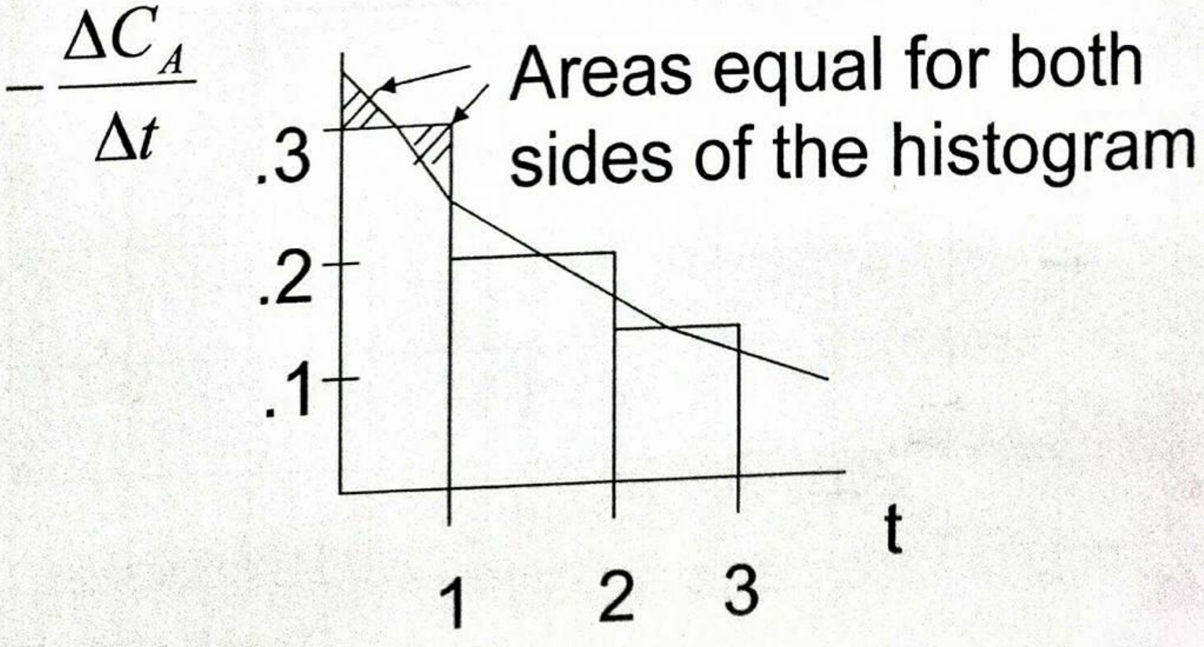
$$\left(\frac{dC_A}{dt}\right)_{i=3} = \frac{0.3 - 0.6}{2(10)} = -0.015$$

Last point:

$$\left(\frac{dC_A}{dt}\right)_{t_5} = \frac{1}{2\Delta t}[C_{A3} - 4C_{A4} + 3C_{A5}]$$

1   0.6   0.4 ← (0.3)

$$\left(\frac{dC_A}{dt}\right)_{i=4} = \frac{6.6) - 4(0.4) + 3(0.3)}{2(10)} = -0.005$$



Areas equal for both sides of the histogram

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(0905421) Chemical Reaction Engineering I	**In-Class Assessment # 19 (Chapter 5)**
Partner (1) Name: زين سلامة	**Partner (2) Name:**
Partner (1) ID #: 0153646	**Partner (2) ID #:**

P5-7 The liquid-phase reaction of methanol and triphenyl takes place in a batch reactor at 25°C

$$\underset{(B)}{CH_3OH} + \underset{(A)}{(C_6H_5)_3CCl} \rightarrow (C_6H_5)_3COCH_3 + HCl$$

For an equal molar feed, the following concentration-time data was obtained for methanol:

$C_A = C_B - (1-y)$

(B)

t (h)	0	0.278	1.389	2.78	8.33	16.66
C_A(mol/dm³)	0.1	0.95	0.816	0.707	0.50	0.37

$-r_A = kC_A^{\alpha} C_B^{\beta}$
$= k\, C_A^{\alpha} C_A^{\beta}$
$= k\, C_A^{(\alpha+\beta)}$

The following concentration-time data was carried out for an initial methanol concentration of 0.01 mol/dm³, and an initial triphenyl of 0.1 mol/dm³:

$C_{A0} = 0.01$ mol/L
$C_{B0} = 0.1$ mol/L
$C_B = C_{B0} =$ constant

(A)

t (h)	0	1	2	5	10
C_A(mol/dm³)	0.01	0.00847	0.00735	0.00526	0.00357

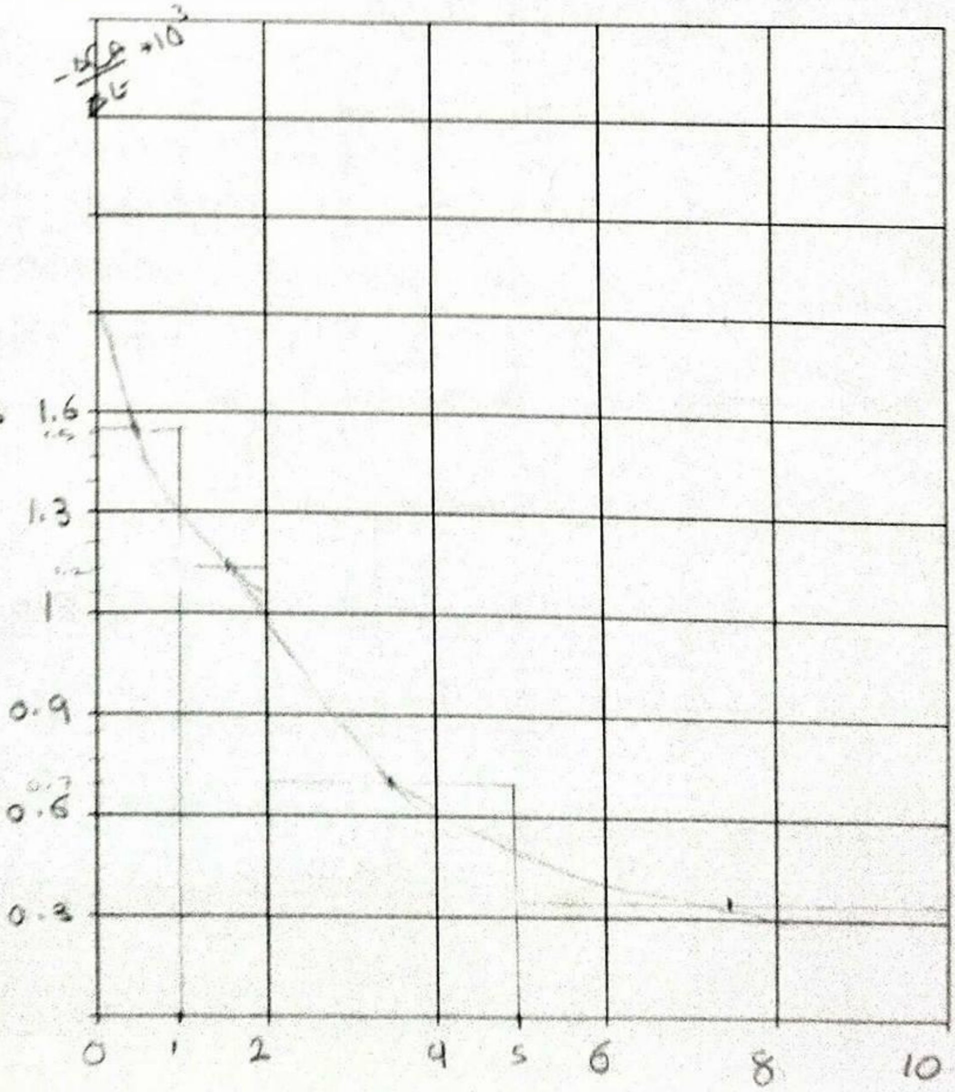
[illegible]
[illegible]

Determine the rate law and rate law parameters.

$-r_A = k' C_A^{\alpha}$

Graphically

t	$-\Delta C_A/\Delta t$
0	
	1.53×10^{-3}
1	
	1.12×10^{-3}
2	
	0.697×10^{-3}
5	
	0.338×10^{-3}
10	



P5-10 The gas-phase decomposition $A \rightarrow B + 2C$ is carried out in a constant volume batch reactor. Runs 1 through 5 were carried out at 100 °C, while run 6 was carried out at 110 °C.
(a) From the data in the table, determine the reaction order and specific reaction rate.
(b) What is the activation energy for this reaction?

α k

E

at 100°C

Run	Initial Concentration, C_{A0} (mol/L)	Half-Life, $t_{1/2}$ (min)	$\ln C_{A0}$	$\ln t_{1/2}$
1	0.0250	4.1	-3.68	1.41
2	0.0133	7.7	-4.3	2.04
3	0.010	9.8	-4.6	2.28
4	0.05	1.96	-2.99	0.67
5	0.075	1.3	-2.6	0.26
6	0.025	2.0	-3.68	0.693

ثابته k → 6
[illegible]
(different temp. at 110°C)

$$\Rightarrow \text{slope} = \frac{1.7-1}{-4.1+(3.4)} = -1$$

$\boxed{a}$

$$\Rightarrow \text{slope} = 1-\alpha$$

$$-1 = 1-\alpha$$

⇒ 2nd order

$\boxed{\alpha = 2}$ order of rxn. (2nd order)

$$\ln t_{1/2} = \ln \frac{2^{\alpha-1}-1}{k(\alpha-1)} + (1-\alpha)\ln C_{A0}$$

$$(0.693) = \ln \frac{2^1-1}{k(1)} + (-1)(-3.68)$$

$\boxed{k_2 = 19.83}$ dm³/mol.min

$\boxed{b}$

⇒ from Arhenus eq:

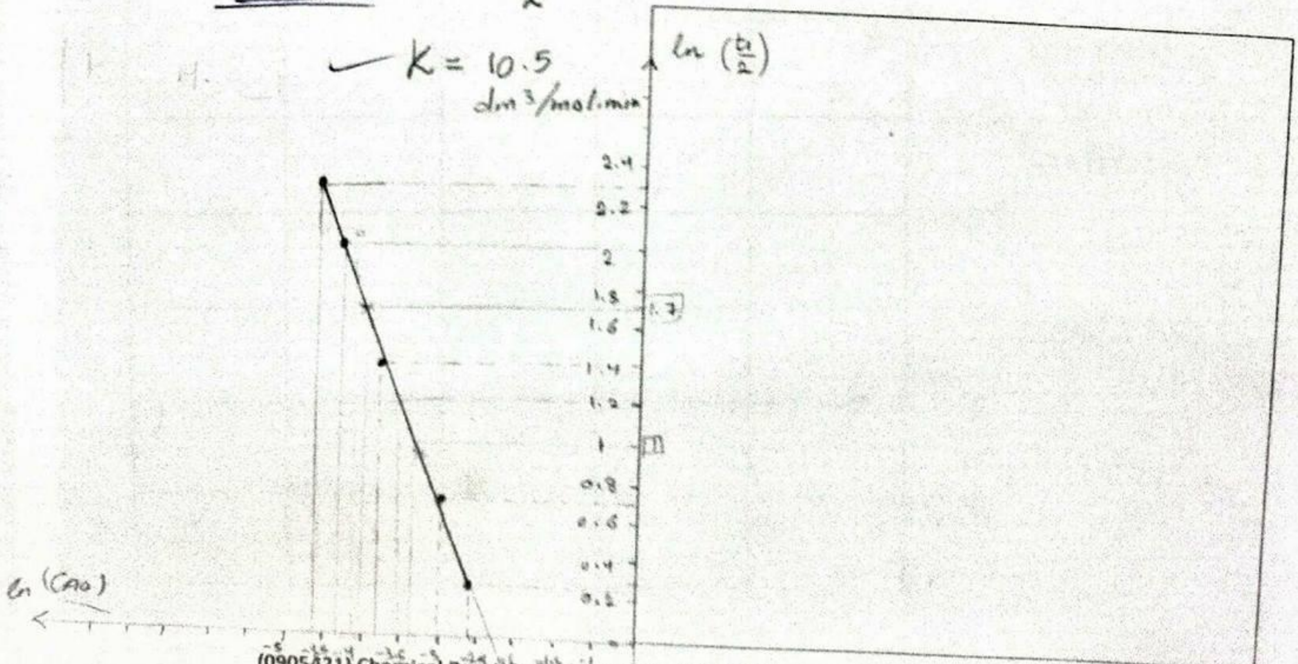
$$\ln\left(\frac{K_1}{K_2}\right) = -\frac{E}{R}\left(\frac{1}{T_1} - \frac{1}{T_2}\right)$$

$$\ln\left(\frac{10.5}{19.83}\right) = \frac{-E}{8.314}\left(\frac{1}{373} - \frac{1}{383}\right)$$

$E = 75518.3$ J/mol

from Intersept ⇒ $\ln \frac{2-1}{k} =$

$K = 10.5$ dm³/mol.min



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Chemical Engineering Department

(0905421) Chemical Reaction Engineering I	**In-Class Assessment # 21 (Chapter 6)**
Partner (1) Name: [illegible]	**Partner (2) Name:** ____________
Partner(1) ID #: ____________	**Partner(2) ID #:** ____________

P6-12 The following liquid-phase reactions were carried out in a CSTR at 325 K.

$3A \longrightarrow B + C$	$-r_{1A} = k_{1A}C_A$	$k_{1A} = 7.0\ min^{-1}$
$2C + A \longrightarrow 3D$	$r_{2D} = k_{2D}C_AC_C^2$	$k_{2D} = 3.0\ L^2/(mol^2.min)$
$4D + 3C \longrightarrow 3E$	$r_{3E} = k_{3E}C_DC_C$	$k_{3E} = 2.0\ L/(mol.min)$

The concentrations measured *inside* the reactor were $C_A = 0.10$, $C_B = 0.93$, $C_C = 0.51$, and $C_D = 0.049$ all in mol/L.

(a) What are r_{1A}, r_{2A}, and r_{3A}?

$\boxed{r_1A = -k_{1D} C_A}$ $\boxed{r_3A = 0}$

$$\frac{r_2A}{-1} = \frac{r_2D}{3}$$

$$\boxed{r_2A = -\frac{k_{2D}}{3} C_A C_C^2}$$

(b) What are r_{1B}, r_{2B}, and r_{3B}?

$\boxed{r_1B = \frac{k_1}{3} C_A}$ → becareful for the Signs

$\boxed{r_2B = 0}$

$\boxed{r_3B = 0}$

(c) What are r_{1C}, r_{2C}, and r_{3C}?

$\boxed{r_1C = \frac{k_1}{3} C_A}$

$\boxed{r_2C = \frac{-2}{3} k_2 C_A C_C^2}$

$\boxed{r_3C = -k_3 C_D C_C}$

(d) What are r_{1D}, r_{2D}, and r_{3D}?

$\boxed{r_1D = 0}$

$\boxed{r_2D = k_2 C_A C_C^2}$

$\boxed{r_3D = \frac{-4}{3} k_1 C_D C_C}$

(e) What are r_{1E}, r_{2E}, and r_{3E}?

$$\boxed{r_{1E} = 0}$$

$$\boxed{r_{2E} = 0}$$

$$\boxed{r_{3E} = k_3 C_D C_C}$$

(f) What are the net rates of formation of A, B, C, D, and E?

$$r_{A,net} = -k_1 C_A - \frac{k_2}{3} C_A C_C^2$$

$$r_{B\text{-}net} = \frac{k_1}{3} C_A$$

$$r_{C,net} = \frac{k_1}{3} C_A - \frac{2}{3} k_2 C_A C_C^2 - k_3 C_D C_C$$

$$r_{D\text{-}net} = k_2 C_A C_C^2 - \frac{4}{3} k_3 C_D C_C$$

$$r_{E\,net} = k_3 C_D C_C$$

(g) The entering volumetric flow rate is $\overset{\upsilon_0}{\boxed{100\text{ L/min}}}$ and, the entering concentration of A is $\underset{C_{A0}}{\boxed{3\text{ M}}}$. What is the CSTR reactor volume?

Parts (h), (j), (j) are HOMEWORK! (Solve them using POLYMATH or MATLAB)

$\upsilon_0 = 100$ L/min

$C_{A0} = 3$ mol/L

$V_{CSTR} = ?$

$C_C = 0.51$ mol/L

$C_A = 0.1$ mol/L

$$V = \frac{F_{A0} - F_A}{-r_{A,net}} = \frac{(C_{A0} - C_A)\upsilon_0}{-r_A} = \frac{F_C - F_{C0}}{r_{A,net}}$$

$$= \frac{(3 - 0.1)100}{k_1 C_A + \frac{k_2}{3} C_A C_C^2}$$

~~= 400 L~~

$$= \frac{(3 - 0.1)100}{(7)(0.1) + \frac{3}{3}(0.1)(0.51)^2}$$

$$\boxed{V_{CSTR} = 399.4\ L}$$

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The desired liquid-phase reaction $A + B \xrightarrow{k_1} \overset{desired}{R} + T \qquad r_R = 1.0C_AC_B^{0.3}$

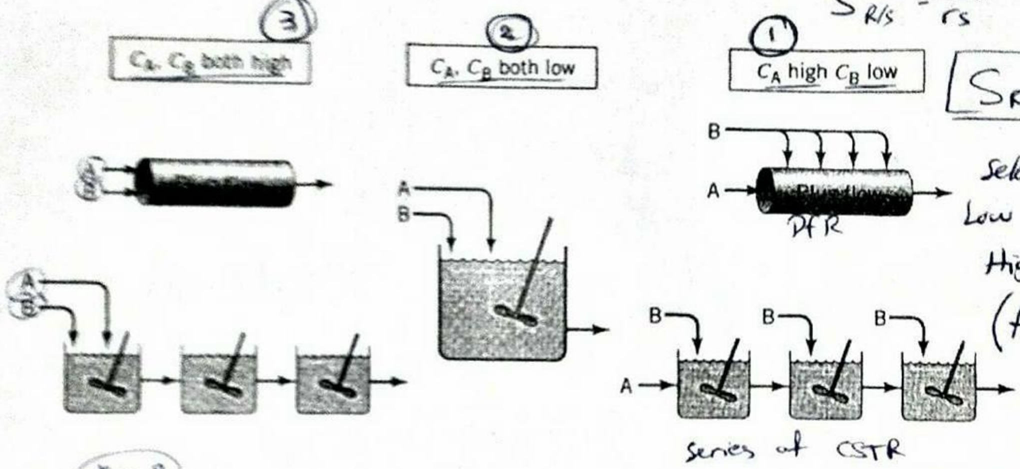
is accompanied by the unwanted side reaction $A + B \xrightarrow{k_2} S + U \qquad r_S = 1.0C_A^{0.5}C_B^{1.8}$

a. From the standpoint of favorable product distribution, order the contacting schemes shown below from the most to the least desirable.

رتب

$$S_{R/S} = \frac{r_R}{r_S} = \frac{C_A\ C_B^{0.3}}{C_A^{0.5}\ C_B^{1.8}}$$

$$\boxed{S_{R/S} = C_A^{0.5}\ C_B^{-1.5}}$$



بناءً على Selectivity
لابد يكون ال Low
ويكون ال High
(High C_A
Low C_B)
if I can't
I Prefer
(low C_A
low C_B)
because the effecte
for $C_B^{-1.5}$ Larger
than $C_A^{0.5}$.

b. For 90% ($X_A = .9$) conversion of pure A and pure B, both of which is fed from a stream of 20 mol/L (C_{A0}) concentration, what will be the overall yield of R, C_R, and C_S in a

1. plug flow reactor, PFR

$C_{A0}' = 20$ mol/h $\quad C_{A0} = 10$ mol/L $\qquad X_A = 90\%$

$C_{B0}' = 20$ mol/h $\quad C_{B0} = 10$ mol/L $\qquad \Rightarrow C_A = 1$ mol/L $= C_B$

1. PFR $\qquad Y_R = \dfrac{F_R}{F_{A0} - F_A} = \dfrac{C_R}{C_{A0} - C_A}$

$dv = \dfrac{dF_A}{r_A} = \dfrac{v_0\, dC_{A0}}{-C_A^{1.3} - C_A^{2.3}}$; $C_A = C_B$

Class work #(21)

ⓑ

$$\left.\begin{array}{l}\longrightarrow C_{A0}' = 20 \text{ mol/L} \\ \longrightarrow C_{B0}' = 20 \text{ mol/L}\end{array}\right] \longrightarrow \begin{array}{l} C_{A0} = 10 \text{ mol/L} \\ C_{B0} = 10 \text{ mol/L}\end{array}$$

$X_A = 0.9$ assume same volume flow rate $\underline{\upsilon_0}$

$$C_A = C_{A0}(1 - X)$$

$$= 10(1 - 0.9)$$

$$\boxed{C_A = 1 \text{ mol/L}} = \underline{\underline{C_B}}$$

Ⓑ PFR: (Plug flow reactor)

overall yeild of R:

$$Y_R = \frac{F_R}{F_{A0} - F_A} = \frac{\upsilon_0 C_R}{\upsilon_0 C_{A0} - \upsilon_0 C_A} = \frac{\overset{??}{C_R}}{\underset{10\swarrow}{C_{A0}} - \underset{1\swarrow}{C_A}} = \frac{1.7}{10-1} = \boxed{0.19}$$

very Low

$\underline{\underline{19\%}}$

⇒ ① mole balance

$$\frac{dF_A}{dV} = r_A$$

$$dV = \frac{dF_A}{r_A} = \frac{\upsilon_0\, dC_A}{-C_A C_B^{0.3} - C_A^{0.5} C_B^{1.8}} = \frac{\upsilon_0\, dC_A}{-C_A^{1.3} - C_A^{2.3}}$$

$\hookrightarrow \boxed{C_A = C_B}$

$$\boxed{dV = \frac{dF_A}{r_A} = \frac{dF_R}{r_R} = \frac{dF_S}{r_S}} \; *$$

I need to know $\underline{\underline{C_R}}$:

$$\frac{dF_A}{r_A} = \frac{dF_R}{r_R} \Rightarrow \frac{\upsilon_0\, dC_A}{-(C_A^{1.3} + C_A^{2.3})} = \frac{\upsilon_0\, dC_R}{C_A^{1.3}} \rightarrow \frac{-dC_A}{\cancel{C_A^{1.3}}(1 + C_A)} = \frac{dC_R}{\cancel{C_A^{1.3}}}$$

$$\int_{C_{A0}}^{C_A} \frac{-dC_A}{1 + C_A} = \int_{C_{R0}=0}^{C_R} dC_R \Rightarrow \int_{C_A=1}^{C_{A0}=10} \frac{dC_A}{1 + C_A} = C_R$$

$$\ln\left(\frac{1 + C_{A0}}{1 + C_A}\right) = C_R \Rightarrow C_R = \ln\left(\frac{1+10}{1+1}\right)$$

$$\boxed{C_R = 1.7 \text{ mol/L}}$$

⇒ because the stoich. are equivalent

$$C_{A0} = C_A + C_R + C_S$$

$$10 = 1 + 7.1 + C_S$$

$$\boxed{C_S = 7.3 \text{ mol/L}}$$

② mixed reactor (CSTR):

① mole balance

$$V = \frac{F_{A0} - F_A}{-r_A} = \frac{F_R - \cancel{F_{R0}}^{0}}{r_R}$$

$$\frac{\cancel{v_0}(C_{A0} - C_A)}{C_A^{1.3} + C_A^{2.3}} = \frac{v_0 C_R}{C_A^{1.3}} \Rightarrow C_R = \frac{C_{A0} - C_A}{(1 + C_A)} = \frac{10 - 1}{1 + 1} =$$

$$\boxed{C_R = 4.5 \text{ mol/L}}$$

$$C_{A0} = C_A + C_R + C_S$$

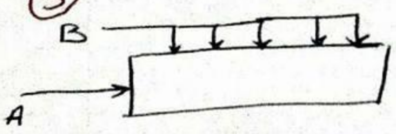
$$10 = 1 + 4.5 + C_S$$

$$\boxed{C_S = 4.5 \text{ mol/L}}$$

⇒ Yeiled of R ⇒ $Y_R = \frac{F_R}{F_{A0} - F_A} = \frac{C_R}{C_{A0} - C_A} = \frac{4.5}{10 - 1} = \boxed{0.5}$

50%

③



$C'_{A0} = \boxed{20}$ mol A/L

$v^{\circ} = 1$ L/min

$C'_{B0} = 1$ mol B/L

$v_0 = 0.05$ L/min

1 stream / 20 streams

total molar flowrate

$$\boxed{C_{A_0} = \frac{v_0 \cancel{C'_{A0}}}{v_{total}} = \frac{(1)(20)}{1 + 0.05} = 19 \frac{\text{mol}}{L}}$$

$C_{A0} = ?$

as base = 1 L/min

ثابت

$$\boxed{C_B = 1 \text{ mol/L}}$$

$$\boxed{C_A = 1 \text{ mol/L}}$$

$C_A \neq C_B$

$$\boxed{r_R = C_A C_B^{0.3} = C_A}$$

$$-r_A = r_R + r_S$$

$$\boxed{-r_A = C_A + C_A^{0.5}}$$

$$\frac{dC_A}{r_A} = \frac{dC_R}{r_R} \rightarrow \int_{C_{A0}}^{C_A} \frac{-dC_A}{C_A + C_A^{0.5}} = \int_0^{C_R} \frac{dC_R}{C_A}$$

$$\int_{C_A=1}^{C_{A0}=19} \frac{dC_A}{1 + C_A^{-0.5}} = C_R \Rightarrow \text{or calculator}$$

~~$\ln\left(\frac{1 + 19^{-0.5}}{1 + 1^{-0.5}}\right) = C_R$~~

$$\boxed{C_R = 13.25 \text{ mol/L}}$$

$$\boxed{Y_R = \frac{13.25}{19 - 1} = 0.74}$$

$$C_S = C_{A0} - C_A - C_R$$

$$= 19 - 1 - 13.3$$

$$\boxed{C_S = 4.7 \text{ mol/L}}$$

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(0905421) Chemical Reaction Engineering I	**In-Class Assessment # 20 (Chapter 6)**
Partner (1) Name: [illegible]	**Partner (2) Name:** ______________
Partner (1) ID #: ______________	**Partner (2) ID #:** ______________

P6-6 Consider the following system of gas-phase reactions:

$A \longrightarrow X \qquad r_X = k_1 C_A^{1/2} \qquad k_1 = 0.004\ (mol/L)^{1/2}/min$

$A \longrightarrow \boxed{B} \qquad r_B = k_2 C_A \qquad k_2 = 0.3\ \text{min}^{-1}$

$A \longrightarrow Y \qquad r_Y = k_3 C_A^2 \qquad k_3 = 0.25\ L/(mol. min)$

color

B is the desired product, and X and Y are foul pollutants that are expensive to get rid of. The given specific reaction rates are at 27 °C. The reaction system is to be operated at 27 °C and 4 atm. Pure A enters the system at a volumetric flow rate of 10 L/min. v_0

(a) Sketch the instantaneous selectivities ($S_{B/X}$, $S_{B/Y}$, and $S_{B/XY} = r_B/(r_X + r_Y)$) as a function of the concentration of A, C_A.
(b) Consider a series of reactors. What should be the volume of the first reactor?
(c) What are the effluent concentrations of A, B, X, and Y from the first reactor?
(d) What is the conversion of A in the first reactor?
(e) If 99% conversion of A is desired, what reaction scheme and reactor sizes should you use to maximize $S_{B/XY}$?
(f) Suppose that E_1 = 20,000 cal/mol, E_2 = 10,000 cal/mol, and E_3 = 30,000 cal/mol. What temperature would you recommend for a single CSTR with a space time of 10 minutes and an entering concentration of A of 0.1 mol/L?

$\boxed{\tau = 10\ min}$

$\boxed{C_{A0} = 0.1}$

$S_{B/X} = \frac{r_B}{r_X} = \frac{k_2 C_A}{k_1 C_A^{1/2}} = \frac{k_2}{k_1} C_A^{1/2}$

$\ln k = \ln A + \frac{ER}{T}$

Low temp.

$\boxed{P6\text{-}6}$

gas phase

$A \rightarrow X$ $\qquad r_X = k_1 C_A^{1/2}$ $\qquad k_1 = 0.004\ (mol/L)^{0.5}/min$

$A \rightarrow \textcircled{B}$ desired $\qquad r_B = k_2 C_A$ $\qquad k_2 = 0.3\ min^{-1}$

$A \rightarrow Y$ $\qquad r_Y = k_3 C_A^2$ $\qquad k_3 = 0.25\ L/mol \cdot min$

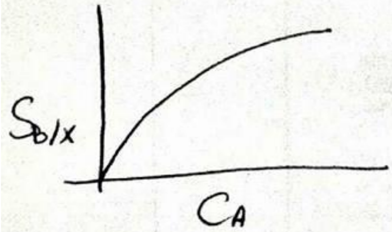
Pure A

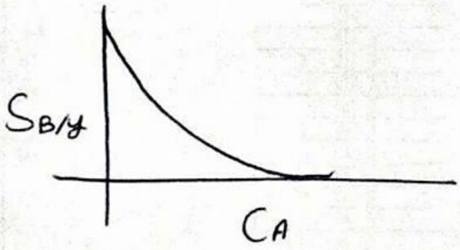
$\upsilon_0 = 10\ L/min$

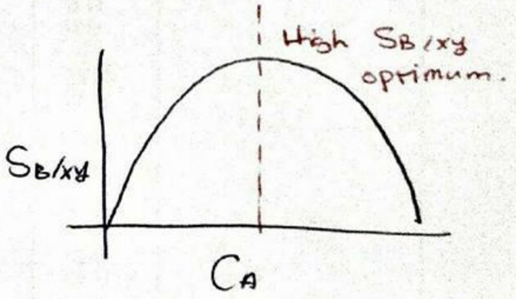
ⓐ $S_{B/X} = \frac{r_B}{r_X} = \frac{k_2 C_A}{k_1 C_A^{0.5}} = \frac{k_2}{k_1} C_A^{0.5} = \frac{0.3}{0.004} C_A^{0.5} = 75\ C_A^{0.5}$

$S_{B/Y} = \frac{r_B}{r_Y} = \frac{k_2 C_A}{k_3 C_A^2} = \frac{0.3}{0.25} C_A^{-1} = 1.2\ C_A^{-1}$

$S_{B/X+Y} = \frac{r_B}{r_X + r_Y} = \frac{k_2 C_A}{k_1 C_A^{0.5} + k_3 C_A^2} =$







$S_{B/XY\,max.} = 10$ at $C_A^* = 0.04$ mol/L

$\Downarrow$

SO, the CSTR should be used with exist conc. C_A^*.

Also, $C_{A0} = \frac{P}{RT} = 0.162 \frac{mol}{L}$

ⓑ

$-r_A = r_X + r_B + r_Y$

$= k_1 C_A^{0.5} + k_2 C_A + k_3 C_A^2$

$V = \frac{\upsilon_0 (C_{A0} - C_A^*)}{-r_A} = \frac{10(0.162 - 0.04)}{k_1 C_A^{*0.5} + k_2 C_A^* + k_3 C_A^{*2}}$

$= \frac{10(0.162 - 0.04)}{(0.004)(0.04)^{0.5} + (0.3)(0.04) + (0.25)(0.04)^2}$

$\boxed{V = 92.4\ L}$

⇒ How find $\underline{C_A^*}$ at max. $S_{B/XY}$:

$$S_{B/XY} = \frac{k_2 C_A}{k_1 C_A^{0.5} + k_3 C_A^2}$$

$$\frac{dS_{B/XY}}{dC_A} = \frac{k_2 (k_1 C_A^{0.5} + k_3 C_A^2) - k_2 C_A (0.5 k_1 C_A^{-0.5} + 2 k_3 C_A)}{k_1 C_A^{0.5} + k_3 C_A^2}$$

$$0 = (0.3)(0.004 * C_A^{0.5} + 0.25\, C_A^2) - (0.3) C_A (0.5 * 0.004\, C_A^{-0.5} + 2(0.25) C_A)$$

↳ $\boxed{C_A^* = 0.04 \text{ mol/L}}$ ⇒ max. selectivity $= 10$

~~$0 = 1.2*10^{-3} C_A^{0.5} + 0.075\, C_A^2 - 6*10^{-4} C_A^{0.5} - 0.15\, C_A^2$~~

~~$0 = 6*10^{-4} C_A^{0.5} + 0.075\, C_A^2 - 0.15\, C_A$~~

ⓒ $\tau = \frac{V}{v_0} = \frac{92}{10} = 9.2$ min

$$V_{CSTR} = \frac{F_B - \cancel{F_{B0}}^{\,0}}{r_B}$$

$$V = \frac{v_0 C_B}{r_B}$$

$$\tau = \frac{C_B}{r_B}$$

$$9.2 = \frac{C_B}{k_2 C_A^*}$$

$$(9.2)(0.3)(0.04) = C_B^*$$

$$\boxed{C_B^* = 0.11 \text{ mol/L}}$$

$$V = \frac{F_X - \cancel{F_{X0}}^{\,0}}{r_X}$$

$$\tau = \frac{C_X}{k_1 C_A^{0.5}}$$

$$C_X^* = (9.2)(0.004)(0.04)^{0.5}$$

$$\boxed{C_X^* = 0.0074 \ \frac{\text{mol}}{\text{L}}}$$

$$V = \frac{F_Y - \cancel{F_{Y0}}^{\,0}}{r_Y}$$

$$\tau = \frac{C_Y}{k_3 C_A^2}$$

$$C_Y^* = (9.2)(0.25)(0.04)^2$$

$$\boxed{C_Y^* = 0.0037 \ \frac{\text{mol}}{\text{L}}}$$

ⓓ conversion of (A) in the 1st reactor

$$X = \frac{C_{A0} - C_A}{C_{A0}} = \frac{0.162 - 0.04}{0.162} = 0.75$$

ⓔ CSTR followed by a PFR should be used.

required $X = 0.99$

$\Rightarrow$ for PFR, mole balance : $\frac{dV}{dX} = \frac{F_{A0}}{-r_A}$

$$\int dV = \int \frac{F_{A0}}{-r_A}\, dx$$

$$= \frac{v_0 C_{A0}}{k_1 C_A^{0.5} + k_2 C_A + k_3 C_A^2} \int_{0.75}^{0.99} dx$$

~~$V = \frac{(10)(0.162)}{(0.004)(0.04)^{0.5} + (0.3)(0.04) + (0.25)(0.04)^2} \int_{0.75}^{0.99} dx$~~

~~$V = (122.73)(0.99 - 0.75)$~~

assume isothermal isobaric operation

and we know $\boxed{\varepsilon = y\delta = 0}$

$\Rightarrow \boxed{v = v_0}$

$\Rightarrow C_A = C_{A0}(1 - x) \Rightarrow C_A = 1.62 * 10^{-3}$

$$V = (10)(0.162) \int_{0.75}^{0.99} \frac{dx}{(0.004)\underset{0.162 *}{C_{A0}^{0.5}}(1-x)^{0.5} + (0.3)(0.162)(1-x) + (0.25)(0.162)^2(1-x)^2}$$

[redacted] [redacted] [redacted] [redacted]

$$\boxed{V = 92\ L}$$

~~$\int^{0.99} dx$~~

(0905421) Chemical Reaction Engineering I In-Class Assessment # 24 (Chapter 8)

Partner (1) Name: زين أحمد [illegible] **Partner (2) Name:** ____________

Partner (1) ID #: ____________ **Partner (2) ID #:** ____________

Example: PBR with Heat Exchange and Pressure Drop

The elementary reversible reaction $A + B \leftrightarrow C$ is carried out in a PBR with constant coolant temperature T_C. We want to determine both the conversion and temperature profiles.

Write a Polymath program to plot X and T vs. W.

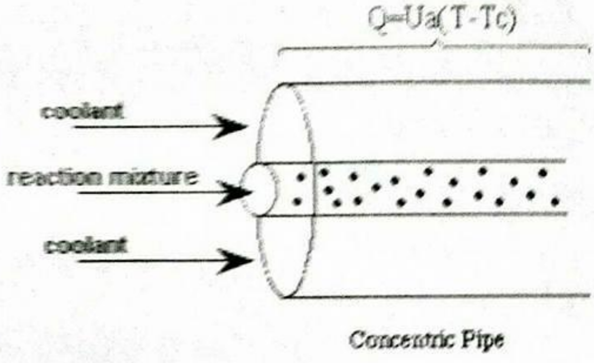
$v_0 = 50$ L/s

$F_{A0} = 5$ mol/s

$F_{B0} = F_{A0}$

$F_I = F_{A0}$

$\boxed{T_o = 325 \text{ K}}$



Concentric Pipe

$E = 25$ kcal/mol

$\Delta H_{Rx} = -20$ kcal/mol

$k = 0.004 \frac{dm^6}{kg \cdot mol \cdot S}$ @310 K

$K_C = 1000$@303 K

$\alpha = 0.0002/$kg

$W = 2500$ kg

$$C_{P_A} = C_{P_B} = C_{P_I} = 20 \text{ cal/mol/K}$$

$$C_{P_C} = 40 \text{ cal/mol/K}$$

$$\frac{Ua}{\rho_b} = 0.5 \frac{cal}{kg \cdot s \cdot K}$$

$$\boxed{T_a = 325 \text{ K}}$$

Solution:

1) Mole Balance:

$$\frac{dX}{dW} = \frac{-r_A'}{F_{A0}} \qquad F_{A0} = 5 \text{ mol/s}$$

2) Rate Law:

$$-r_A' = k\left(C_A C_B - \frac{C_C}{K_c}\right)$$

$$k = k_1 \exp\left(\frac{-E}{R}\left(\frac{1}{T} - \frac{1}{T_1}\right)\right) \qquad k_1 = 0.004 \text{ dm}^3/\text{kmol.s}$$

$T_1 = 310$ K

$E = 25{,}000$ cal/mol , $R = 1.987 \frac{cal}{mol.K}$

$$K_c = K_{c2} \exp\left(-\frac{\Delta H_{Rx}}{R}\left(\frac{1}{T} - \frac{1}{T_1}\right)\right) \qquad K_{c2} = 1000$$

$T_2 = 303$ K

$\Delta H_{Rx} = 20{,}000$ cal/mol

$\boxed{\Delta C_p = 0}$ يعني

$\downarrow$

لأن DH_{Rx}

ثابتة ما بتعتمد T

$$C_A = \frac{C_{A0}(1-x)}{(1+\epsilon x)} y \frac{T_0}{T}$$

$$C_B = \frac{C_{A0}(\Theta_B - x)}{1+\epsilon x} y \frac{T_0}{T}$$

$$C_C = \frac{C_{A0} X}{1+\epsilon x} y \frac{T_0}{T}$$

$C_{A0} = \frac{5}{50} = 0.1$ mol/L

$y_{A0} = \frac{F_{A0}}{F_{T0}} = \frac{F_{A0}}{F_{A0}+F_{B0}+F_{I0}}$

$= \frac{F_{A0}}{3F_{A0}} = \frac{1}{3}$

$\varepsilon = y_{A0}\delta = \frac{1}{3}(-1) = -\frac{1}{3}$

$T_0 = 325\ k$

$\frac{dy}{dw} = \frac{-\alpha}{2y}(1+\varepsilon x)\frac{T_0}{T}$ NOT const Temp. $\varepsilon \neq 0$

$\alpha = 0.0002\ kg^{-1}$

For a packed bed reactor ($dw = \rho_b\, dv$) where $\underline{\rho_b}$ = density at $\underline{\underline{\text{bulk}}}$

The Energy Balance for PBR Reactor:

$$\boxed{\frac{dT}{dw} = \frac{\frac{Ua}{\rho_b}(T_a - T) + r_A' \Delta H_{RX}}{F_{A0}(\Sigma\Theta_i C_{pi} + \Delta C_p X)}}$$

$\frac{Ua}{\rho_b} = 0.5 \frac{cal}{kg.s.k}$

$T_a = 325\ K$

$\Sigma\Theta_i C_{pi} = C_{pA} + \Theta_B C_{pB} + \Theta_I C_{pI}$

$= 20 + (1)(20) + (1)(40)$

$= 80 \frac{cal}{mol.k}$

ⓐ $W = 0$, $X = 0$, $y = 1$

$T = 325\ k$

$W_F = 2500\ kg$

by Polymath solve $\boxed{\underline{x}}$

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(0905421) Chemical Reaction Engineering I	**In-Class Assessment # 25 (Chapter 8)**
Partner (1) Name: [illegible]	**Partner (2) Name:**
Identity (1) #:	**Identity (2) #:**

P8-5 CSTR with Heat Effect

The endothermic liquid-phase elementary reaction $A + B \rightarrow 2C$ proceeds, substantially, to completion in a single steam-jacketed, continuous-stirred reactor. From the following data, **calculate the steady-state reactor temperature**:

Saturation [illegible]
عندما [illegible] steam [illegible]

Reactor volume: 125 gal   Steam jacket area: 10 ft^2 Agitator shaft horsepower: 25 hp
Jacket steam: 150 psig (365.9 °F saturation temperature)
Overall heat-transfer coefficient of jacket, U: 150 Btu/ h.ft^2. °F
Heat of reaction, ΔH^o_{Rx} = +20,000 Btu/lb mol of A (independent of temperature) $\Rightarrow \Delta C_p = 0$

Component	A	B	C
Feed (lbmol/hr)	10.0	10.0	0
Feed temperature (°F)	80	80	-
Specific heat (Btu/lb mol. °F)	51.0	44.0	47.5
Molecular weight	128	94	222
Density (lb/ft^3)	63.0	67.2	65.0

$$X_{EB} = \frac{\frac{UA}{F_{A0}}(T - T_a) + \Sigma\Theta C_{pi}(T - T_0)}{-\Delta H_{Rx}}$$

$$\Rightarrow 1 = \frac{\frac{150 \text{ Btu}}{h.ft^2.°F}\,\frac{10\ ft^2.hr}{10 \text{ lbmol}}(T - 365.9)°F + \frac{-63525 \text{ Btu}/h}{10 \text{ lbmol}/h} + 95(T - 80)}{W}$$

But there is W_s

$\dot{Q} = UA(T_a - T)$

$\dot{Q} - \dot{W}_s \Rightarrow$ NOT Neglection for shaft work

$\boxed{\dot{W}_s = -25 \text{ hp}} = -63525$ Btu/hr
done on the system.

$\dot{Q} = \boxed{UA(T - T_a) + \dot{W}_s}$

$\Sigma\Theta_i C_{pi} = C_{pA} + (1) C_{pB}$
$= 95 \frac{Btu}{lbmol.F}$ ✓

[illegible] k

-20000 Btu/lbmol

Solution:

$V = \frac{F_{A0}X}{-r_A}$

[illegible]

$-r_A = k C_A C_B$

$C_A = C_{A0}(1 - X)$
$C_B = C_{A0}(1 - X)$

$\Big\} \; -r_A = k C_{A0}^2 (1 - X)^2$

$k = k_1 \exp\left(\frac{-E}{R}\left(\frac{1}{T} - \frac{1}{T_1}\right)\right)$

$C_p = \Sigma$ Prod. $- \Sigma$ reactants

$\boxed{\Delta C_p = 0}$

The University of Jordan
School of Engineering
Chemical Engineering Department

(0905421) Chemical Reaction Engineering I	**In-Class Assessment # 23 (Chapter 8)**
Partner (1) Name: رغد أحمد البيروتي	**Partner (2) Name:**
Partner (1) ID #:	**Partner (2) ID #:**

Temp. change

P8-6 The elementary irreversible organic liquid-phase reaction $A + B \rightarrow C$ is carried out **adiabatically** in a flow reactor. An equal molar feed in A and B enters at 27°C, and the volumetric flow rate is 2 dm³/s and $C_{A0} = 0.1\ kmol/m^3$.

const v

ν_0

$$f_{A0} = \nu_0 C_{A0} = \frac{2\ dm^3}{s}\left|\frac{0.1\ kmol}{m^3}\right|\frac{1\ m^3}{10^9\ dm^3} = \frac{0.2\times10^{-9}\ kmol}{s}$$

Additional information:

OH reference Temp.

$H_A^o(273\ K) = -20\ kcal/mol$ | $C_{p_A} = C_{p_B} = 15\ cal/(mol.K)$ | $k = 0.01\ L/(mol.s)$ at $300\ K$

$H_B^o(273\ K) = -15\ kcal/mol$ | $C_{p_C} = 30\ cal/(mol.K)$ | $E = 10{,}000\ cal/mol$

$H_C^o(273\ K) = -41\ kcal/mol$

$V = f(T)$ [illegible]
$x = f(T)$

(a) Calculate the PFR and CSTR volumes necessary to achieve 85% conversion. What are the reasons for the differences?

$R = 1.987\ \frac{cal}{mol.K}$

Note: *You can use 'Simpson's $\frac{3}{8}$ rule" if numerical integration is needed:*

$$\int_{X_0}^{X_3} f(x)dx = \frac{3h}{8}[f_0 + 3f_1 + 3f_2 + f_3]$$

$$h = \frac{X_3 - X_0}{3}$$

CSTR:

liquid ✓
elementry $A+B \rightarrow C$ ✓
adiabatic ✓
$F_{A0} = F_{B0}$
$T_0 = 27°C = 300\ K$
$\nu_0 = 2\ dm^3/s$
$C_{A0} = 0.1\ kmol/m^3$
$= 0.1\ mol/dm^3$
$x = 85\%$

① — $V = f_{A0}\int_0^{0.85} \frac{dx}{k_1 \exp\left(\frac{-E}{R}\left(\frac{1}{T} - \frac{1}{T_1}\right)\right) C_{A0}^2 (1-x)^2}$

$* \ \Delta C_p = C_{p_C} - C_{p_B} - C_{p_A}$
$= 30 - 15 - 15 = 0$

energy bal → $\Rightarrow T = T_0 + \frac{-\Delta H_{Rx} \cdot X}{\sum \Theta_i C_{p_i}}$

ⓐ PFR

mole balance : $V = F_{A0}\int_0^x \frac{dX}{-r_A}$

rate Law : $-r_A = k C_A C_B$

$k = k_1 \exp\left(-\frac{E}{R}\left(\frac{1}{T} - \frac{1}{T_1}\right)\right)$

$k_1 = 0.01\ L/mol.s$ @ $T_1 = 300\ K$

stoich: $C_A = C_{A0}(1-x)$
$C_B = C_{A0}(\Theta_B + \nu_B x)$
$= C_{A0}(1-x)$ $\Rightarrow -r_A = k C_A^2 (1-x)^2$

$\Delta H_{Rx} = H_C - H_B - H_A$
$= -6\ kcal/mol$
$= -6000\ cal/mol$

$T = 300 + \frac{6000\ x}{C_{p_A} + \Theta_B C_{p_B}} = \frac{6000\ X}{15 + (1)(15)} + 300$

$T = 300 + \frac{6000}{30} x$

$\boxed{T = 300 + 200\ X}$

① 3 [illegible]

exothermic ↑T ↑x

(b) What is the maximum inlet temperature one could have so that the boiling point of the liquid (550 K) would not be exceeded even for complete conversion?

$\rightarrow (X=1)$

$T = T_0 + 200X$ $\qquad T_0 = 350\ K$

$\overset{\Vert}{550} \qquad \overset{\Vert}{1}$

(d) Calculate the conversion that can be achieved in one 500-dm³ CSTR and in two 250-dm³ CSTRs in series.

$X = ?$ $\qquad V_{CSTR} = 500\ dm^3$

$$V = \frac{F_{A0} X}{k C_{A0}^2 (1-X)^2}$$

$k \rightarrow k_1 \exp\left(-\frac{E}{R}\left(\frac{1}{T} - \frac{1}{T_1}\right)\right)$

$k_1 \rightarrow 0.01$, $T \rightarrow 300+200x$, $T_1 \rightarrow 300$

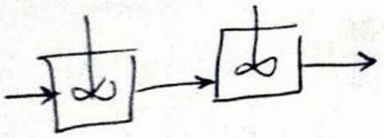
by trial & error

$X = 0.5$, $V = 60\ L$

$X = 0.9$, $V = 333\ L$

$X = 0.93$, $V = 618\ L$

$X = 0.921$, $V = 499\ L$



$\boxed{0.87 \rightarrow 217L}$

$$V_1 = 250 = \frac{0.2 X_1}{k(0.1)^2 (1-X_1)^2}$$

$X_1 = 0.882$

$X = 0.7 \Rightarrow V = 74.8$

$X = 0.8 \Rightarrow V = 116.9$

$X = 0.9 \Rightarrow V = 338L$

$V = 250\ L \Rightarrow X_1 = 0.862$

$$250 = V_2 = \frac{F_{A0}(X_2 - 0.882)}{k C_{A0}^2 (1-X_2)^2}$$

$0.9 - 0.882 = 0.018$

$X_2 = 0.945 \rightarrow V_2 = 134\ L$

$X_2 = 0.98 \rightarrow V_2 = 600\ L$

$X_2 = 0.967$

(0905421) Chemical Reaction Engineering I — In-Class Assessment # 26 (Chapter 8)

Partner (1) Name: معتز عبد الباري — Partner (2) Name: ____________

Identity (1) #: ____________ — Identity (2) #: ____________

The first-order irreversible exothermic liquid-phase reaction A→ B is to be carried out in a jacketed CSTR. Species A and an inert I are fed to the reactor in equilmolar amounts. The molar feed rate of A is 80 mol/min. The related G(T)-R(T) and ignition-extinction curve are shown below.

Additional information:

Heat capacity of the inert: 30 cal/mol.°C

$\tau = 100$ min

Heat capacity of A and B: 20 cal/mol.°C

$\Delta H_{Rx} = -7500$ cal/mol

UA = 8000 cal/min.°C

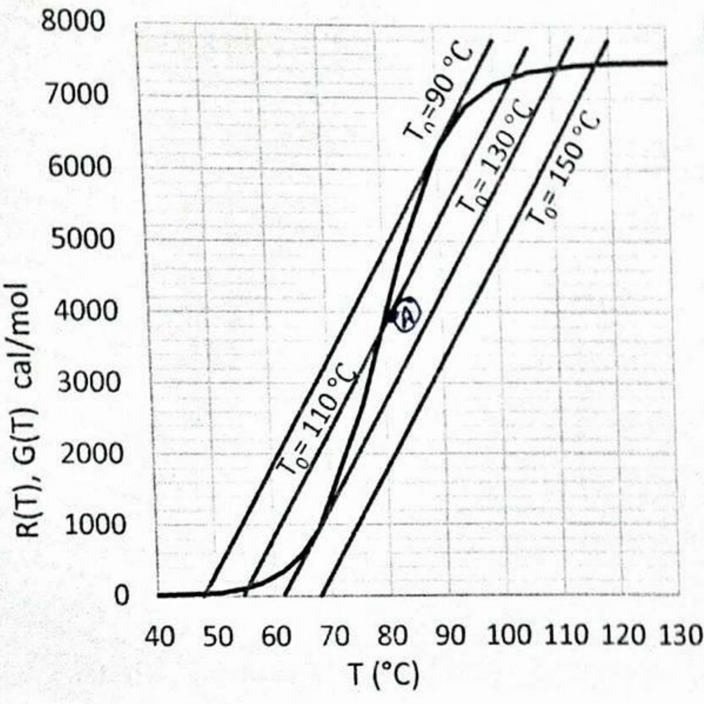
$k = 6.6\times10^{-3}\ \text{min}^{-1}$ at 350 K

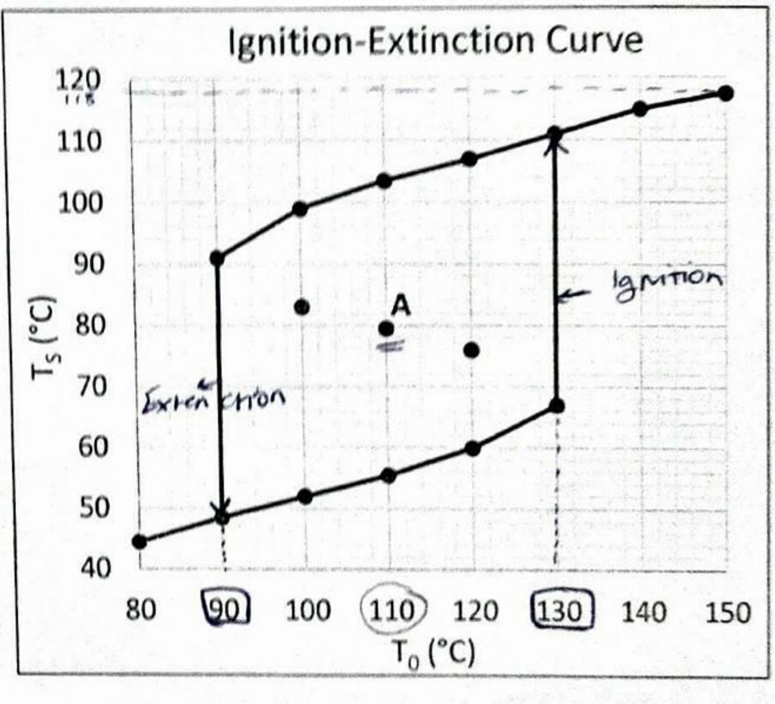
E = 40 kcal/mol.K

Ambient temperature, $T_a = 300$ K

(a) What is the extinction and ignition temperatures of this system?

T(extinction) = 90 °C — T (ignition) = 130 °C





(b) Mark point Ⓐ on the ignition-extinction curve back on G(T)-R(T) curve.
Is this a locally stable steady state point? Why? NO, It's ~~[illegible]~~ ~~Un-steady stable state~~ an Un-stable, becaus do not return to it if there is a temp. Pulse.

(c) What is the reactor temperature for a feed temperature of 150 °C? 118 °C

(d) For the reactor to operate at a high conversion, the feed fluid must be preheated to at least 130 °C. Explain this statement.

(e) Suppose that the fluid is now heated to 140 °C and then cooled to 120 °C, where it remains. What will be the reactor temperature and conversion?

School of Engineering
Chemical Engineering Department

(0905421) Chemical Reaction Engineering I	**In-Class Assessment # 27 (Chapter 8)**
Partner (1) Name: رغد البدوي	**Partner (2) Name:**
Identity (1) #:	**Identity (2) #:**

P8-13 The reaction $A + B \Longleftrightarrow C + D$ is carried out adiabatically in a series of staged FBR packed-bed reactors with inter-stage cooling. The lowest temperature to which the reactant stream may be cooled is 27 °C. The feed is equal molar in A and B and the catalyst weight in each reactor *is* sufficient to achieve 99.9% of the equilibrium conversion. The feed enters at 27 °C. If four reactors and three coolers are available, what conversion may be achieved?

L.T = 27 + 273 = 300 K
$X_e = 0.999$

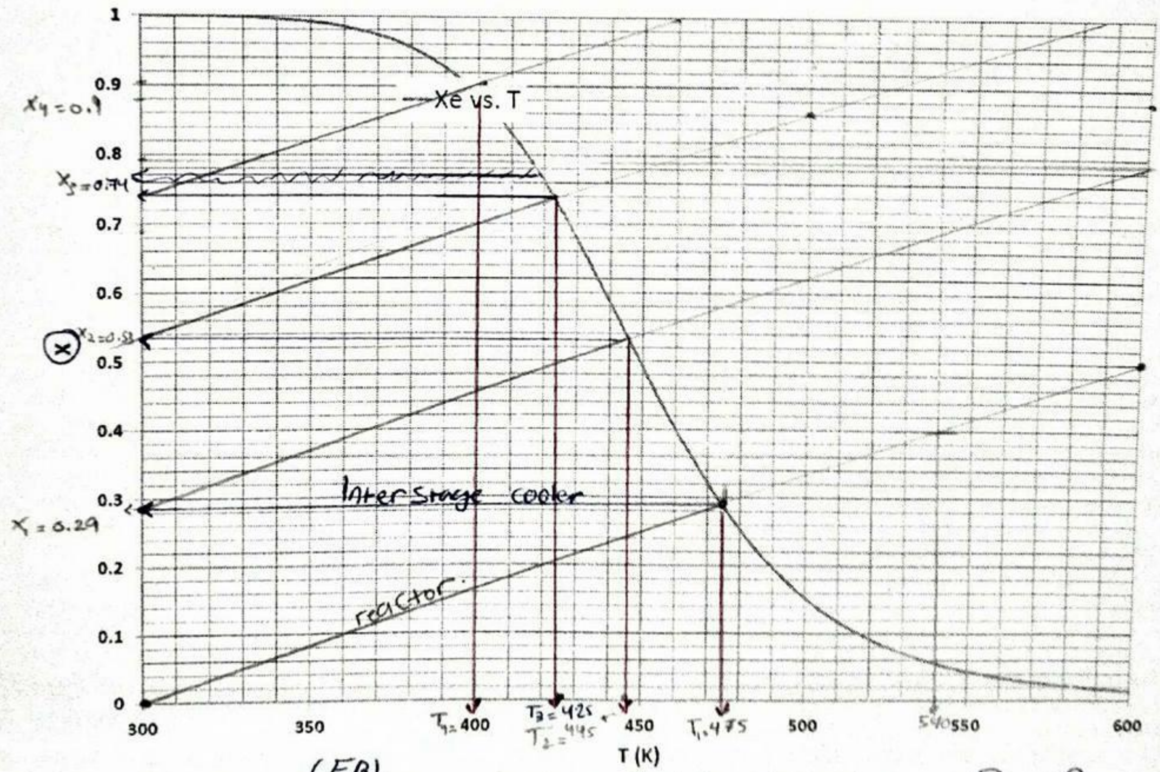
Additional information:

$\Delta H_{Rx} = -30,000\ cal/mol\ A \qquad C_{P_A} = C_{P_B} = C_{P_C} = C_{P_D} = 25\ cal/mol.K$

$K_e\,(50\ °C) = 50,000 \qquad F_{A0} = 10\ mol\ A/min$

Solution:

① Prepare a plot of equilibrium conversion as a function of temperature.



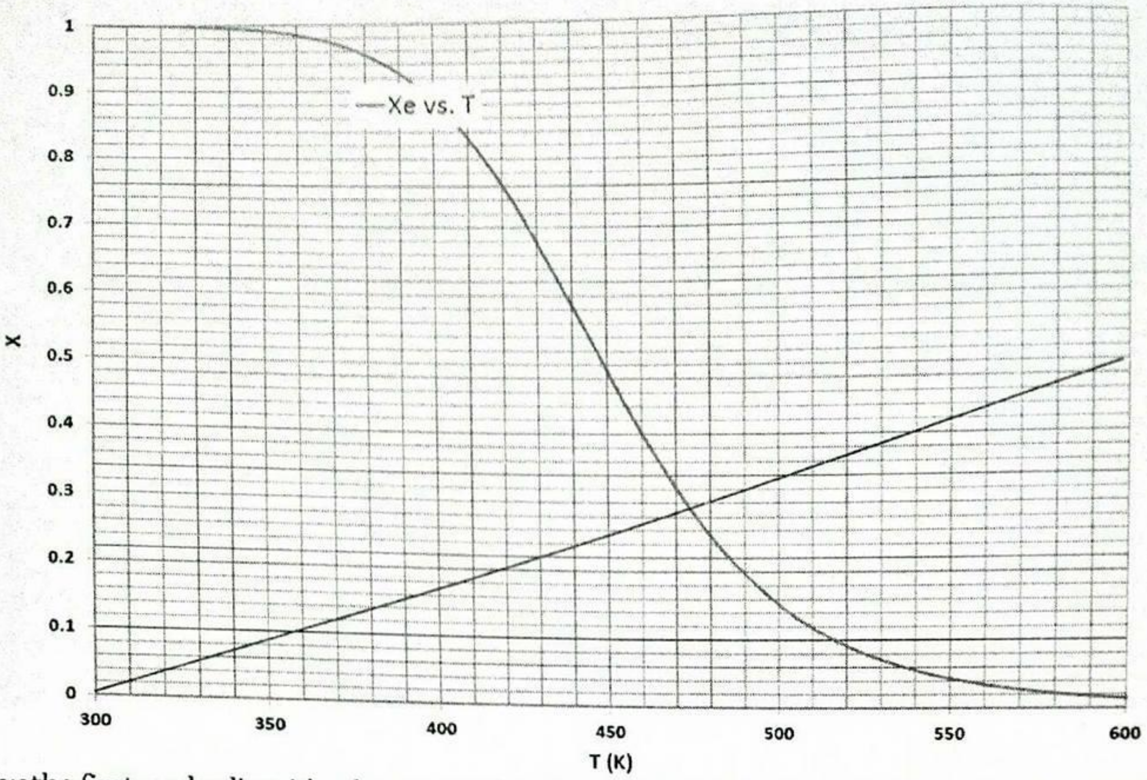
② Use the adiabatic energy balance (EB) equation to find the relation between T and X.

$$T = T_0 + \frac{-\Delta H_{Rx}\cdot X}{\sum \theta_i C_{P_i}} = 300 + \frac{30,000X}{C_{P_A} + C_{P_B}} \rightarrow \boxed{T = 300 + 600X} \Rightarrow (T-300) = 600(X-0)$$

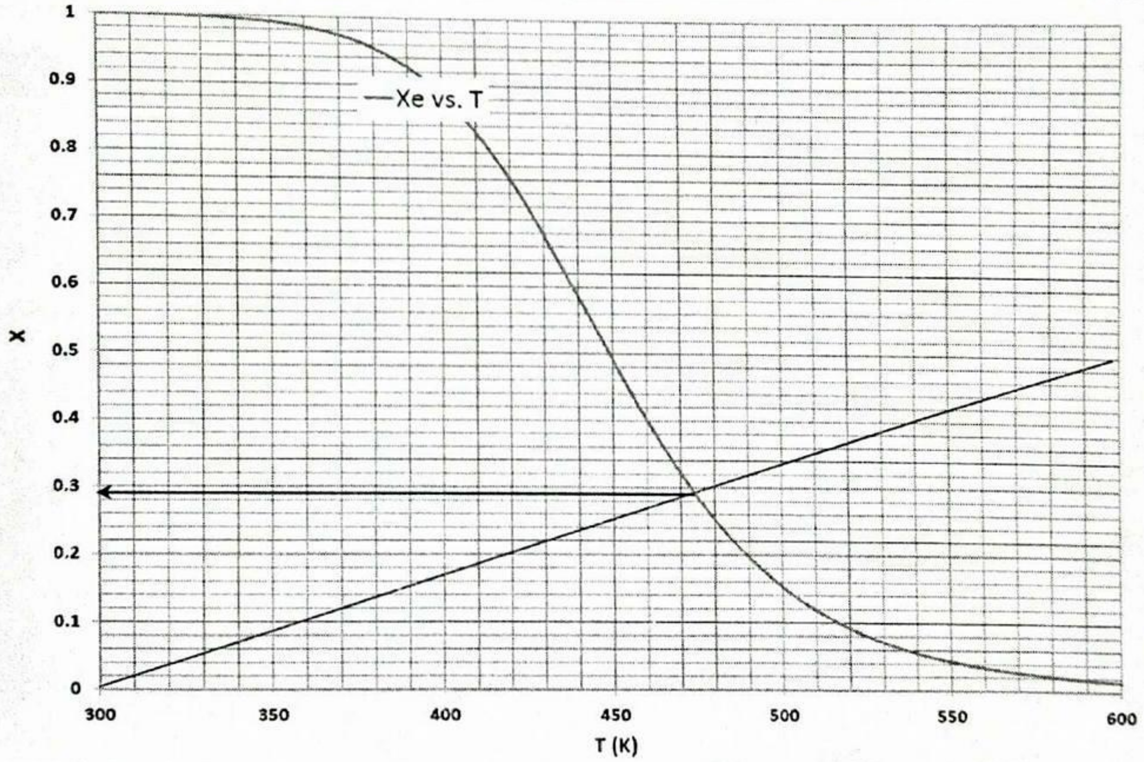
③ Draw the line that represents the first reactor on the equilibrium plot:
(It's a straight line that passes through the 2 points (300, 0) and (600, 0.5))

$X = 0 \Rightarrow T = 300$
$X = 0.5 \Rightarrow T = 600$

From the plot below, we see that $\boxed{X_1 = 0.29}$ and $\boxed{T_1 = 475\ K}$



④ Draw the first cooler line, it's a horizontal line from the first reactor exit temperature to 300 K.

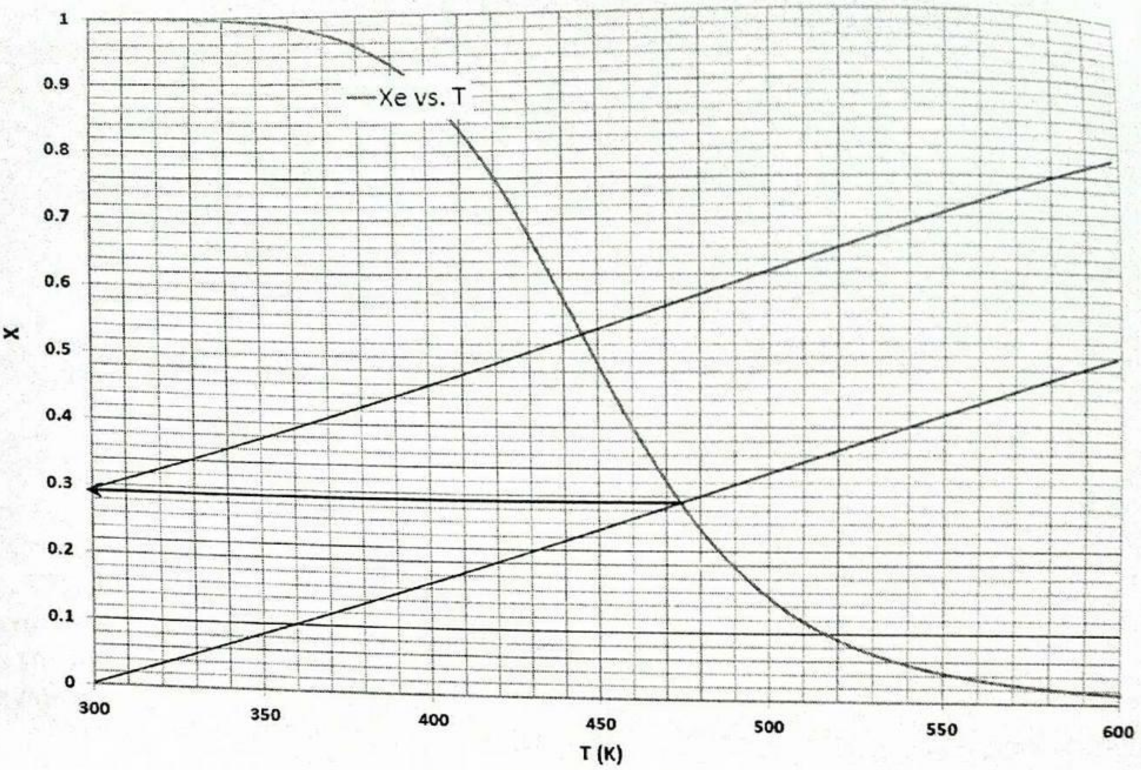


⑤ Draw the line that represents the second reactor on the equilibrium plot; it's parallel to the first reactor but starts from the point (300, 0.29).
You can also draw it by using the adiabatic equation $T = 300 + 600X$ but after remembering that this equation is originally $(T - 300) = 600(X - 0)$, so now it can be used after being modified to $\boxed{(T_2 - 300) = 600(X_2 - X_1)}$

⇨ *(It's a straight line that passes through the 2 points: (300, 0.29) and (600, 0.79)*

From the plot below, we see that $\boxed{X_2 = 0.53}$ and $\boxed{T_2 = 445\ K.}$

$\Rightarrow (600 - 300) = 600\,(x_2 - 0.29) \Rightarrow x_2 = 0.79$



6- Now repeat steps 4 and 5 two times: draw the second cooler horizontal line, the third reactor adiabatic line ($\boxed{X_3 = 0.74}$, and $\boxed{T_3 = 425\ K}$), the third cooler horizontal line, and the fourth reactor adiabatic line ($\boxed{X_4 = 0.90}$, and $\boxed{T_4 = 400\ K}$).