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Krishna Mohan[a]; Rakesh Govind[a]
[a] Department of Chemical Engineering, University of Cincinnati, Cincinnati, Ohio

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STUDIES ON A MEMBRANE REACTOR

Krishna Mohan and Rakesh Govind*
Department of Chemical Engineering
University of Cincinnati
Cincinnati, Ohio 45221

ABSTRACT

Simulation is used to evaluate the performance of a catalytic reactor with permeable wall (membrane reactor) in shifting the equilibrium of three reversible reactions (cyclohexane dehydrogenation, hydrogen iodide decomposition, and propylene disproportionation). It is found that the preferred choice of cocurrent or countercurrent operation is dependent on the physical properties and operating conditions. Methods of enhancing conversion are suggested and temperature effects are discussed.

INTRODUCTION

The concept of shifting the thermodynamic equilibrium in reversible reactions by separating products from the reaction zone is well known. The catalyst is packed on one side of the membrane with products preferentially permeating to the other side, thereby enhancing conversion. This has significant impact on process economics for reactions with unfavorable equilibria.

The objective of this paper is to compare the behavior of a membrane reactor for different reversible reactions of industrial interest. Process variables (design parameters, operating variables, physical properties), and flow patterns (cocurrent, countercurrent, recycle) are used to evaluate performance.

LITERATURE REVIEW

One of the earliest applications of membranes to upset equilibrium was developed by Wood (1). He showed that by forcing non-equilibrium conditions on a hydrogen-porous palladium-silver alloy membrane, an otherwise stable cyclohexane vapor is readily dehydrogenated to cyclohexene. Subsequently, a number of researchers suggested the use of membranes to remove reaction products and thereby shift

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* To whom correspondence should be addressed.

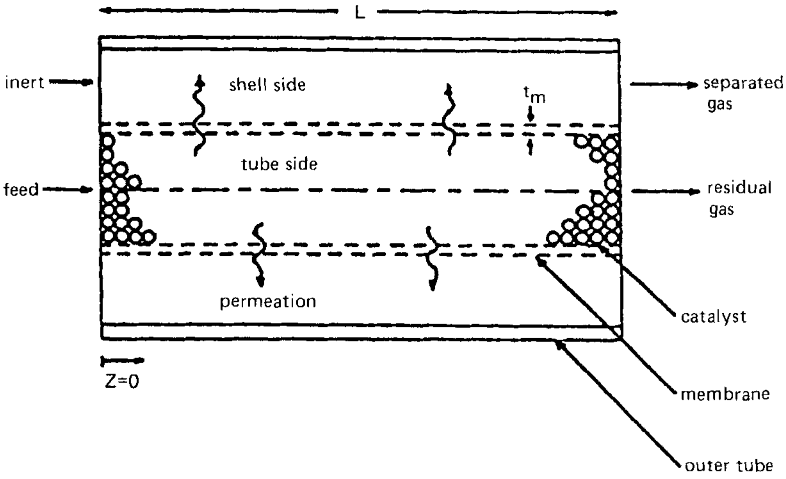


Figure 1. Schematic of membrane reactor

theequilibrium of reversible reactions (2,3,4). Gryaznov et al. (5) conducted hydrogenation and dehydrogenation reactions at opposite surfaces of a membrane catalyst which is selectively permeable to the hydrogen formed at one membrane surface and consumed at the other.

Raymont (6) suggested abundantly available hydrogen sulfide as a possible hydrogen source, and discussed the possibility of using platinum or palladium alloy membranes to shift the equilibrium of the thermodynamically unfavorable decomposition reaction. Since then, a number of experimental studies have shown the advantages of separating hydrogen directly from the reaction zone (7,8,9,10). Since palladium alloy membranes degrade in the presence of hydrogen sulfide above 500°C, most of the studies have used porous Vycor glass membranes. Kameyama et. al. (10) showed the possibility of using porous alumina membranes, which have 30-fold higher permeability than microporous Vycor glass membrane.

In other work, also of significance to thermochemical energy conversion and the 'hydrogen economy', Shinji et. al. (11), and Ito (12) studied cyclohexane dehydrogenation in porous Vycor glass and palladium membranes respectively. Since palladium is permeable only to the hydrogen, high conversions could be achieved by increasing residence time in the membrane reactor. In a separate study, Shindo et. al. (13) employed porous Vycor glass to selectively separate the hydrogen produced by the decomposition of hydrogen iodide.

Theoretical studies of the membrane reactor have shown that there is an optimum thickness of membrane at which maximum conversion is obtained (14,15,16), and when membranes are permeable to reactants, there is a limit to the extent of conversion that can be obtained (14,16).

MODEL DEVELOPMENT

In Figure 1 a schematic of the membrane reactor is shown. Reaction occurs on the side in which catalyst is packed (tube side - ceramic membranes; shell side - hollow fiber membranes), and products preferentially permeate to the low pressure side.

Table 1. Model Equations

$$e_P = \left(\frac{1}{T_r + 1}\right)^{0.5} \quad \text{temperature correction for permeability}$$

$$e_T = \exp\left[\frac{E_A}{RT_0}\left\{1 - \frac{1}{T_r + 1}\right\}\right] \quad \text{temperature correction for reaction rate}$$

$$g_i = \delta\alpha_i(x_i - y_i P_r)Da \qquad \text{permeation flow rate}$$

Reaction side , Z>0 (– for products + for reactants)

$$h_{1i} = \frac{dF_i}{dZ} = \pm f_r e_T - g_i e_P \qquad \text{material balance} \tag{1}$$

$$\frac{dT_r}{dZ} = \left[e_T \Phi(-f_r) - T_r e_P \sum_{i=1}^{N} g_i - \beta(T_r - T_p) - T_r \sum_{i=1}^{N} h_{1i}) \right] / \left[\sum_{i=1}^{N} F_i \right] \tag{2}$$

Energy balance:
heat generation - heat loss (permeation + membrane conduction + convection)

Permeation side, Z>0 (+ for cocurrent flow – for countercurrent flow)

$$h_{2i} = \frac{dQ_i}{dZ} = \pm g_i e_P \tag{3}$$

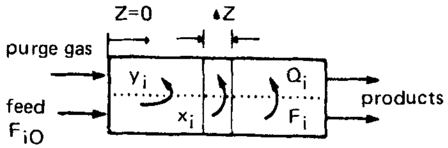
$$\frac{dT_p}{dZ} = \left[\pm\left\{ e_P T_r \sum_{i=1}^{N} g_i + \beta(T_r - T_p) - \psi(T_p - T_a)\right\} - T_p \sum_{i=1}^{N} h_{2i}) \right] / \left[\sum_{i=1}^{N} Q_i \right] \tag{4}$$

$i = 1,2,.....N$ where N is the total number of components (reactants and products)

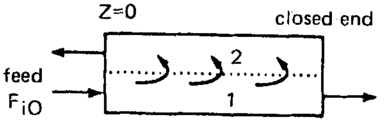
The model equations governing the membrane reactor are essentially differential material and energy balances on either side of membrane. The number of equations obtained depend on the number of components of the gas mixture. For a single reactant decomposing into two products there will be three material balances and one energy balance on either side of membrane. Table 1 lists the general equations governing a non-isothermal membrane reactor.

In this study the following simplifying assumptions have been made:

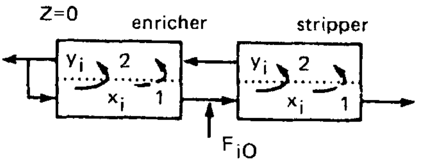
1. One-dimensional plug flow;
2. Negligible axial diffusion flux of heat and mass;
3. Negligible radial gradients of temperature and concentration;
4. Permeability coefficients for gases depend only on membrane characteristics and temperature and are independent of the coupling effects encountered in gas mixtures;



a) cocurrent membrane reactor



b) countercurrent membrane reactor



c) membrane reactor with recycle and intermediate feed location
1 reaction side
2 permeation side

Figure 2. Flow patterns in membrane reactors

5. Isothermal operation - unless otherwise specified;
6. Negligible pressure drop on either side of membrane;
7. Reaction occurs only on the side in which the catalyst is packed.

The design parameters (membrane characteristics and dimensions, reactor dimensions), operating variables (reaction side pressure, permeation side pressure, feed flow rate, inert flow rate, feed temperature), physical properties (reaction rate constant, equilibrium constant, permeabilites) are characterized by dimensionless groups. The Damkohler number, Da, is a measure of the maximum forward reaction rate and is proportional to the length of the membrane reactor. Thus an increase in 'Da' signifies an increase in reaction volume and membrane area. The rate ratio, δ, is inversely proportional to the thickness of membrane, and measures the ratio of the maximum permeation rate of the fastest gas to the maximum reaction rate. The pressure ratio, Pr, is the ratio of the permeation side pressure to the reaction side pressure. The permselectivity ,α_i, is the ratio of the permeation rate of component 'i' to the permeation rate of the fastest gas. Non-isothermal performance is evaluated using the parameter, ΔE, which measures the amount of heat added per mole of reactant converted. The conversion ratio, X_R, (conversion in membrane reactor to equilibrium conversion) is the main index used to measure performance of the membrane reactor.

NUMERICAL SOLUTION

The solution strategy is based on the flow pattern encountered in the membrane reactor (Figure 2). For cocurrent flow the equations are solved as an initial value problem since all the conditions are known at the feed end. In the absence of a purge gas, initial conditions of the permeation side mole fractions are assumed to be the composition of the permeating stream at the membrane reactor feed inlet. The ordinary differential equations are integrated using Adams variable order predictor-corrector method (17).

When the flow is countercurrent a split boundary value problem is obtained, since the necessary conditions to solve the differential equations are known at the feed end on the reaction side and at the opposite end on the permeation side. The iterative technique developed by Fan (18) has been used to adjust the unknown initial conditions. In the past, this method has been used to simulate membrane separators by Antonson et. al.(19), Teslik (20) and Teslik and Sirkar (21).

When there is recycle, successive substitution (22) is used to converge on the compositions of the recycle stream. Similar strategy is utilized when the feed location is not at one end but somewhere in between. Initial guesses are made of the flow rates to the stripper, which is then simulated yielding feed to the enricher. The enricher is a recycle membrane reactor, which on solution yields new guesses of feed composition to the stripper.

The details of the numerical technique employed can be found in Mohan and Govind (23).

REACTIONS STUDIED

The three reactions chosen for this study are cyclohexane dehydrogenation, hydrogen iodide decomposition, and propylene disproportionation. The permeabilities and operating conditions are shown in Table 2 ; kinetic information in Table 3.

The cyclohexane reaction is a typical example of a highly endothermic dehydrogenation reaction. Since most of these reactions need high temperatures to obtain reasonable conversions there may be great economic advantages to using membrane reactors. Applications of this reaction are in thermochemical energy conversion (24), and in hydrogen storage and transmission (25). Rate expressions for the reaction are taken from Ito (12).

Hydrogen iodide decomposition is the reaction which evolves hydrogen in the magnesium-iodine thermochemical water splitting cycle (16). The overall efficiency of the cycle can be improved by enhancing the low equilibrium conversion of reaction. In the absence of available information, permselectivities are based on Knudsen diffusion and typical operating conditions are based on the theoretical study by Ito et. al. (16). Kinetic information is taken from Shindo et. al. (26)

Propylene disproportionation is another reaction which can benefit by the use of a membrane reactor. The Phillips Triolefin Process has been commercially used to produce polymerization grade ethylene and high purity butene from propylene (27). Typical reactor temperatures exceed 500^{O}K, and, using membranes, it will be possible to save considerable energy. This reaction is particularly interesting since it is reversible at room temperatures. Thus, polymer membranes in hollow fiber form can be used to provide high permeation rates. In this study the range of parameters considered encompasses both the tubular Vycor glass reactor and the hollow fiber reactor.

This reaction is being studied currently by Chang and Hwang (28) in a porous Vycor glass reactor. Their permeability studies have indicated that, surface flow is an important transport mechanism and permselectivity cannot be approximated through molecular size, as in Knudsen diffusion. The rate expressions used for this reaction are taken from Kapteijn (29).

RESULTS AND DISCUSSION

In Figure 3 the conversion ratio, X_R, is plotted against the rate ratio, δ, for the three reactions. It can be seen that there is an optimum rate ratio beyond which the conversion decreases. Since the membrane is permeable to all the gases, increasing

Table 2. Reactions Studied

$A \rightleftharpoons B + C$	MEMBRANE	TYPICAL OPERATING CONDITIONS
$C_6H_{12} \rightleftharpoons C_6H_6 + 3H_2$ $\Delta H = 53520$ cal/gmole	VYCOR GLASS $\alpha_A = 0.274$, $\alpha_B = 0.451$ $\alpha_{inert} = 0.224$	$T_0 = 477K$, $K_E = 4.2X10^{-4}$ $\delta = 1.0$, $Da = 20$, $Pr = 1.0$ $F_{inert} = 5.0$, $Q_{inert} = 50.0$
$2HI \rightleftharpoons I_2 + H_2$ $\Delta H = 1400$ cal/gmole	VYCOR GLASS $\alpha_A = 0.125$, $\alpha_B = 0.089$ $\alpha_{inert} = 0.333$	$T_0 = 700K$, $K_E = 0.133$ $\delta = 0.2 - 1.0$, $Da = 10$ $Pr = 0.0$, $F_{inert} = 1.0 - 3.5$
$2C_3H_6 \rightleftharpoons C_2H_4 + C_4H_8$ $\Delta H = 500$ cal/gmole	a) VYCOR GLASS $\alpha_A = 0.43$, $\alpha_B = 0.247$	$T_0 = 298K$, $K_E = 0.065$ $\delta = 0.5$, $Da = 10$ $Pr = 0.0$
	b) SILICONE RUBBER $\alpha_A = 0.4$, $\alpha_B = 0.1$	$T_0 = 298K$, $K_E = 0.065$ $\delta = 3.0$, $Da = 10$ $Pr = 0.0 - 0.5$

Table 3. Kinetic Information

$C_6H_{12} \rightleftharpoons C_6H_6 + 3H_2$
$\Delta H = 53520$ cal/gmole; ($T_0 = 477$)
catalyst: Pt/alumina
$K_E = 4.2X10^{-4} \exp(-\Delta H(1/T - 1/477)/0.48)$; $E_A = 8500$ (cal/mole)
$k_B = 2.03X10^{-10} \exp(6270/T)$

$$f_r = -Da \frac{\left| x_A - \frac{(x_B x_C)}{K_E} \right|}{(x_C^3/K_E + x_A P_h k_B)} \tag{5}$$

$2HI \rightleftharpoons H_2 + I_2$
$\Delta H = 1400$ cal/gmole: ($T_0 = 700K$)
catalyst: Pt/alumina
$K_E = 0.133 \exp(-\Delta H(1/T - 1/700)/0.48)$; $E_A = 6500$ (cal/mole)
$k_B = 1.24X10^{-12} \exp(11522/T)$ (Pa^{-1})

$$f_r = -Da \frac{\left| x_A - \frac{(x_B k_B)^{0.5}}{K_E} \right|}{(1 + (x_B k_B)^{0.5})^2} \tag{6}$$

$2C_3H_6 \rightleftharpoons C_2H_4 + C_4H_8$
$\Delta H = 500$ cal/gmole; ($T_0 = 298K$) negligible temperature effects
catalyst: Re_2O_7/alumina
$K_E = 0.065$; $k_A = 3.3$ $(MPa)^{-1}$; $k_B = 0.72$; $k_C = 5.0$; $k_{AB} = 2.38$ $(MPa)^{-1}$;
$k_{AC} = 17.0$ $(MPa)^{-1}$

$$f_r = -Da(1. + k_A P_h) \frac{\left| x_A^2 - \frac{(x_B x_C)}{K_E} \right|}{(x_A + x_B k_B + x_B k_B + P_h(x_A^2 k_A + k_{AB} x_B x_C + k_{AC} x_A x_C)} \tag{7}$$

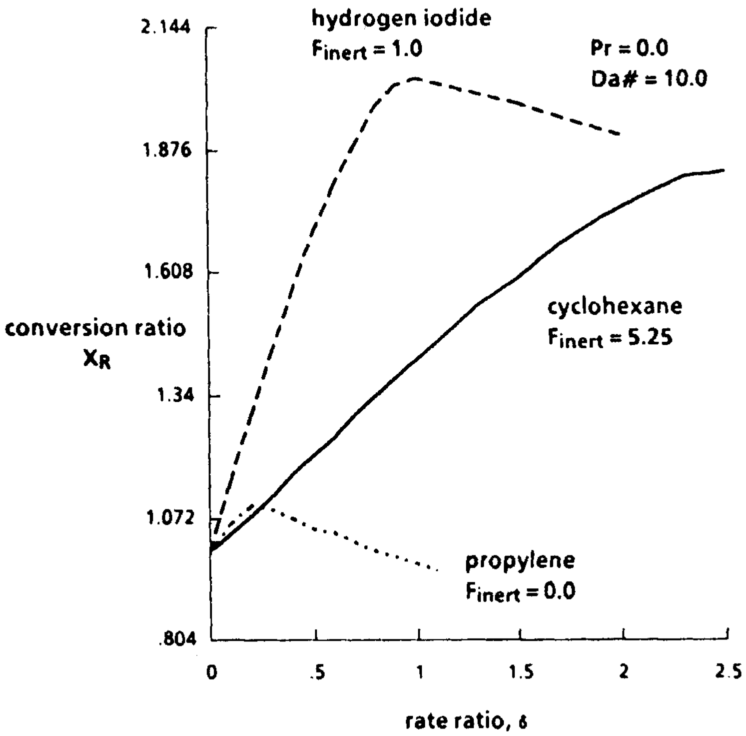


Figure 3. Effect of rate ratio on equilbrium shift in the absence of backpermeation

permeation rate results in considerable loss of reactant. This loss eventually negates the effect of equilibrium shift and there is a decrease in conversion. Reactant loss is particularly severe in the disproportionation reaction due to the high permeability of reactant. Not only is the extent of equilibrium shift quite low but the optimum rate ratio is very small. Thus it is apparent that hollow fiber configuration cannot be used without some modification to the basic membrane reactor.

In Figure 4 (Mohan and Govind, 17), the volume ratio, V_R, is plotted against maximum conversion ratio, X_{RO}, for the cyclohexane reaction. X_{RO} is defined as the ratio of maximum conversion (conversion obtained when the forward reaction rate goes to zero), and the equilibrium conversion. V_R is defined as the ratio of the reaction side volume of a membrane reactor necessary to attain X_{RO}, to the volume of a tubular reactor necessary to attain equilibrium conversion. It should be noted that increasing pressure ratio decreases the equilibrium shift as a result of low permeation rates. Increasing rate ratios also decreases equilibrium shift due to greater loss of reactant. It can also be seen that there is a limiting conversion ratio X_{RM} that can be attained in the membrane reactor.

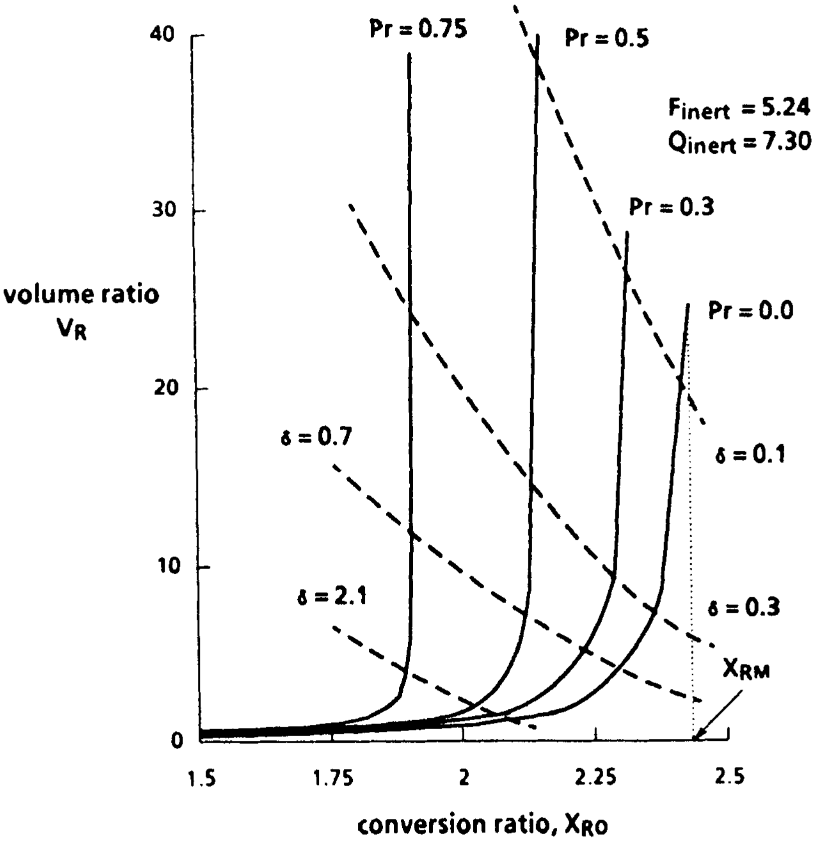


Figure 4. Volume ratio vs. maximum conversion ratio for various rate ratios at different pressure ratios for the cyclohexane dehydrogenation reaction

The limiting conversion ratio is dependent on the operating pressure ratio and on the permselectivity of the reactant. Table 4 lists the limiting conversion at $Pr = 0.0$ for the three reactions.

Cocurrent and Countercurrent Modes of Operation

In designing membrane separators, countercurrent operation is preferred since it provides the larger concentration gradient. However, in membrane reactor design, conditions at the feed end are very important since most of the reaction occurs in a small region near that end. In Figure 5 the difference in conversion ratios between cocurrent and countercurrent, ΔX_R, is plotted the against rate ratio, δ, for cyclohexane dehydrogenation and propylene disproportionation.

Table 4. Maximum Conversion in Membrane Reactor

reaction	conversion		
	X_E	X_{max}	X_R
cyclohexane	0.25	0.60	2.39
hydrogen iodide	0.21	0.45	2.15
propylene (vycor glass)	0.34	0.42	1.235
propylene (hollow fiber)	0.34	0.40	1.12

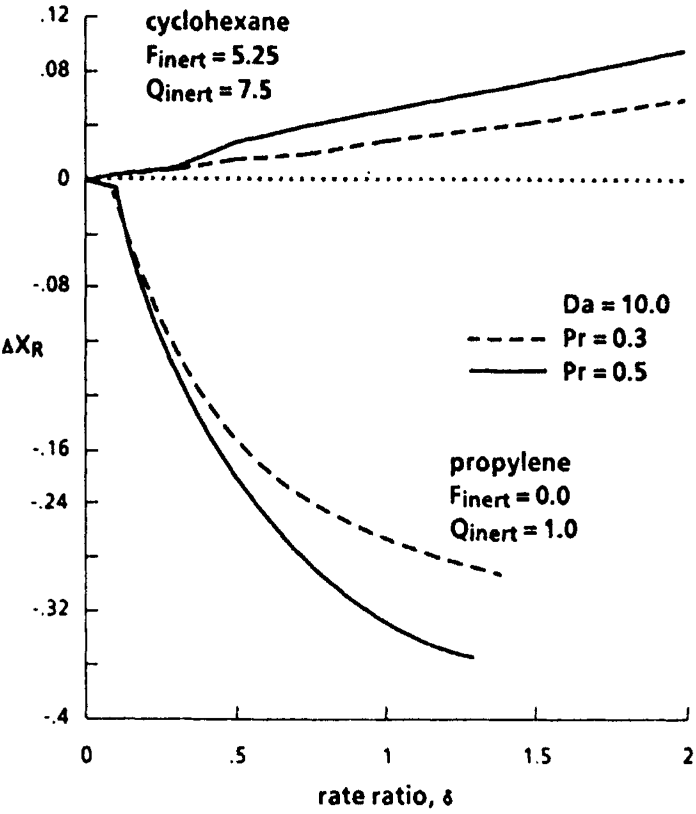


Figure 5. Effect of flow type (cocurrent or countercurrent) on membrane reactor performance for cyclohexane dehydrogenation and propylene disproportionation

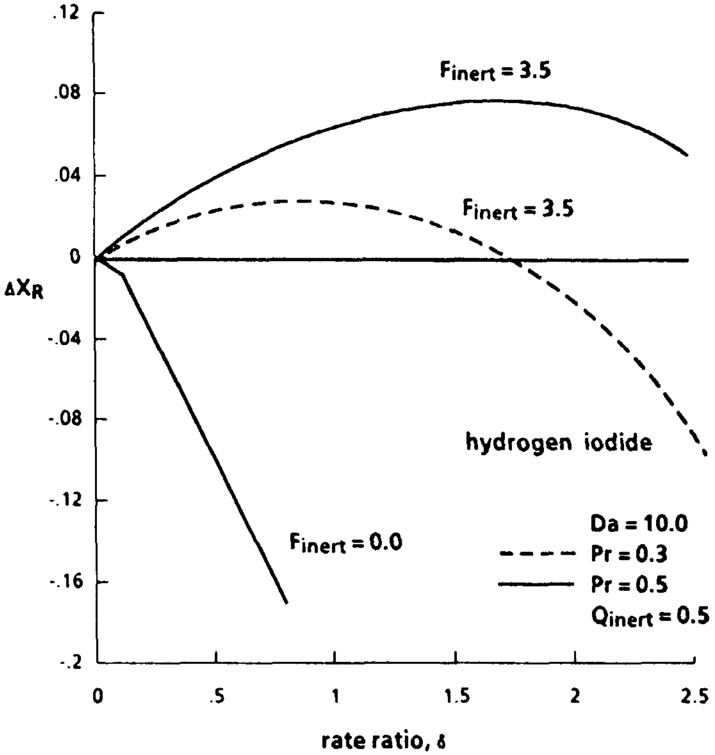


Figure 6. **Effect of flow type on membrane reactor performance for hydrogen iodide decomposition**

For the cyclohexane reaction, cocurrent operation performs better because of the greater equilibrium shift obtained at the feed end. Countercurrent operation has high product concentration on the permeation side at the feed end ; hydrogen permeation rates are low, and as a result forward reaction rates suffer. At high rate ratios and pressure ratios the situation worsens for countercurrent operation as appreciable backpermeation of hydrogen occurs at the inlet.

The opposite scenario is encountered with the propylene disproportionation reaction. Since reactant permeability is high, the controlling factor here is the loss of reactant. Countercurrent operation has high reactant concentrations on the permeation side at the feed end. Reactant permeation rates are low, and as a result, the loss of reactant is less than that for a cocurrent membrane reactor. With increase in rate ratio and pressure ratio, countercurrent operation performance gets significantly better.

For the hydrogen iodide reaction, either factor (equilibrium shift or reactant loss) can control depending on the operating conditions in the membrane reactor, as shown in Figure 6. Thus there is an optimum rate ratio beyond which reactant loss

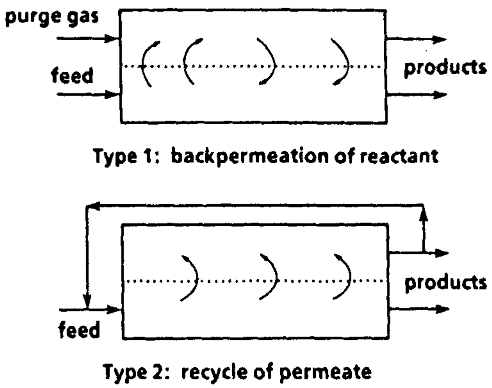


Figure 7. Methods of improving equilibrium shift

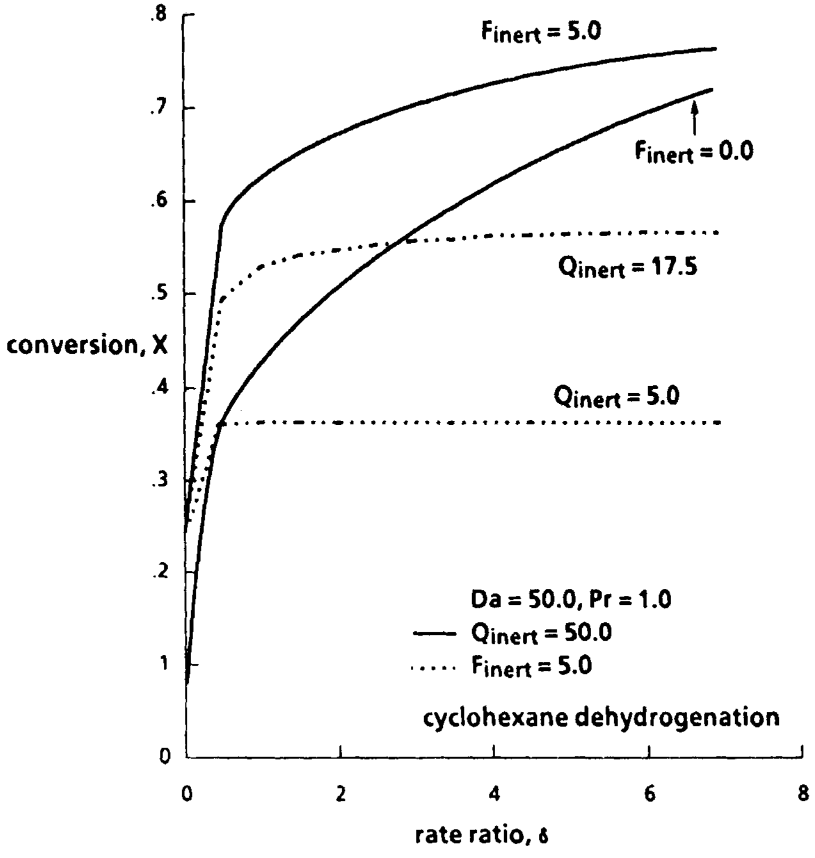


Figure 8. Effect of inerts on membrane reactor performance for cyclohexane dehydrogenation

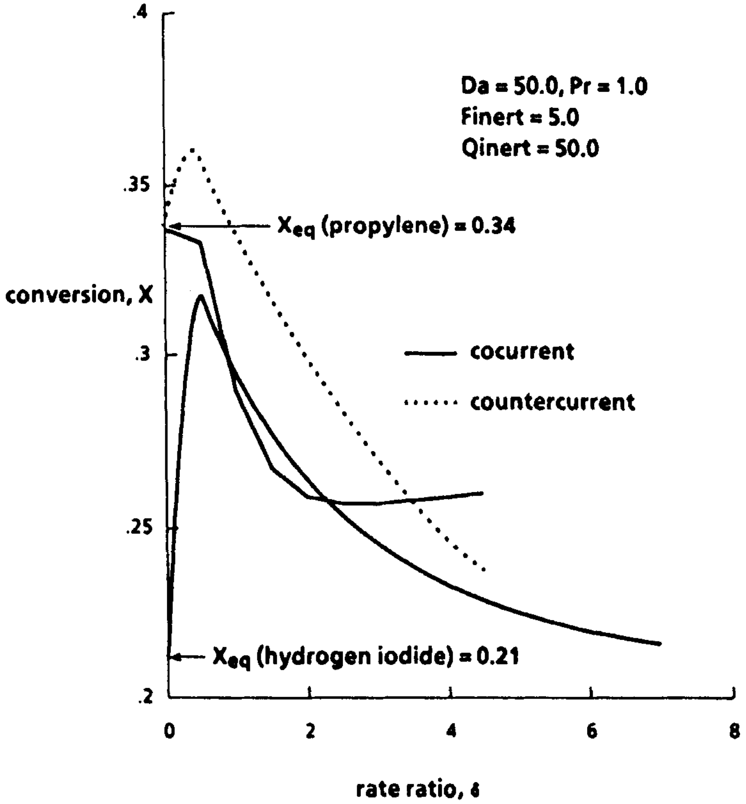


Figure 9. Effect of inerts on membrane reactor performance for hydrogen iodide decomposition and propylene disproportionation

controls and countercurrent operation performs better. When there is no inert on the reaction side there is no optimum since reactant loss controls.

Enhancing Equilibrium Shift

The only way of improving conversion ratio beyond the maximum, X_{RM}, is by reducing the amount of reactant lost to the permeate side. The two possible ways of doing this are shown in Figure 7. Both modes are described below.

Backpermeation. It is obvious that for this mode to be advantageous, the membrane reactor has to be operated cocurrently, and at a pressure ratio of unity. In Figure 8 conversion, X, is plotted against rate ratio for the cyclohexane reaction. Figure

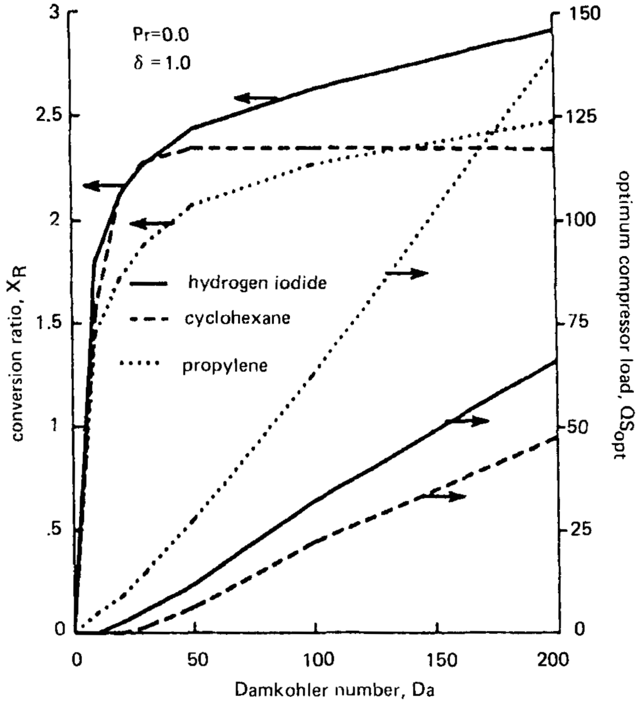
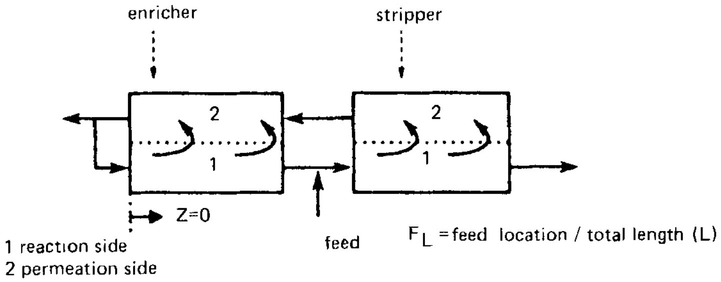


Figure 10. Effect of permeate recycle on membrane reactor performance

9 is a similar plot for the hydrogen iodide and propylene reactions. For the cyclohexane reaction, with a large flow rate on the permeation side, it is possible to achieve high conversions. However, for the other two reactions, this mode of enhancing equilibrium is not very useful. The main reason for this is that cyclohexane dehydrogenation is a reaction in which there is a volume increase with reaction, and it is possible to create conditions where reactant is replenished by backpermeation as it reacts. At low inert flow rates on the permeation side, backpermeation of hydrogen kills any possibility of equilibrium shift. This mode of operation is used by Shinji et. al. (11), in their experimental study of cyclohexane dehydrogenation. To achieve high conversions they used high flow rates on the permeation side (i.e a large amount of inerts) and low flow rates (i.e. large residence times) on the reaction side.

Recycle. In this mode of operation it is necessary to provide external energy to compress the recycle stream. In Figure 10 conversion ratio, X_R, and optimum compressor load, Qs_{opt}, is plotted against the Damkohler number, Da. It can be seen that both hydrogen iodide and propylene reactions can be enhanced by the use of recycle. For both



Pr=0.0

CONVERSION

	δ=1.0		δ=3.0	
Da #	F_L=0.0	F_L=opt	F_L=0.0	F_L=opt
4.0	0.39	0.41	0.47	0.49
10.0	0.49	0.52	0.605	0.62
20.0	0.585	0.61	0.70	0.71
50.0	0.7	0.715	0.8	0.81

Figure 11. Intermediate feed location for propylene disproportionation reaction

reactions since the permeability of reactant is in between that of the products, separation of products can be used to enhance conversions beyond the maximum, X_{max}. There is an optimum compressor load because products are also being recycled and eventually will cause enough backreaction to negate the effects of equilibrium shift. High conversions can be obtained by increasing reaction volume, at the expense of large membrane areas for permeation and energy to compress the recycle stream. The hollow fiber configuration of the membrane reactor is now viable for the propylene reaction.

When permeability of reactant is less than that of the products, product separation cannot be used to increase conversion. Thus for the cyclohexane reaction, the limiting conversion, X_{max}, cannot be exceeded by recycle.

Equilibrium shift can be further enhanced for the propylene disproportionation reaction by shifting the feed location, as shown in Figure 11. Thus the membrane reactor has the same configuration as the Continuous Membrane Column (30) and is divided into two parts: an enriching section in which the ethylene is separated from butene, and a stripping section in which feed is introduced. There is an optimum feed location since, beyond a certain size of enriching section, reactant lost in the bottom stream cannot offset gains in separation.

Temperature Effects

All three reactions studied in this paper are endothermic. However, for the hydrogen iodide and propylene reactions, heat effects are very mild and conversion decrease is small. Cyclohexane dehydrogenation, on the other hand, is highly endothermic and a temperature drop of $40^{O}K$ decreases the equilibrium constant by a

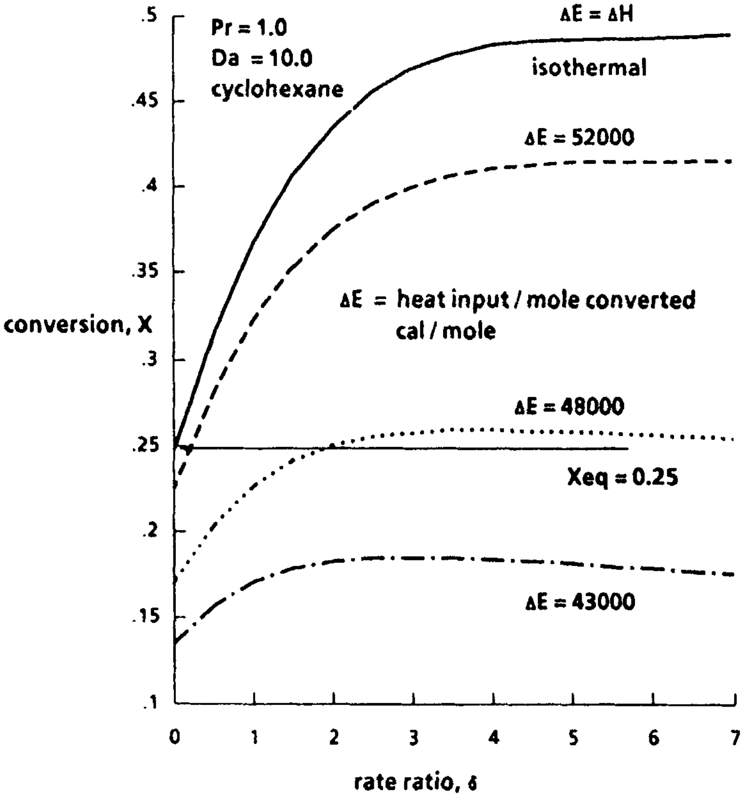


Figure 12. Effect of heat removal on performance of membrane reactor for cyclohexane dehydrogenation

factor of a hundred. Thus, for this reaction, non-isothermal conditions have a significant effect on conversion.

External heat has to be supplied to achieve high conversions. A large amount of inerts should not be used for this purpose since it will decrease permeation rates and mask the effect of equilibrium shift. In Figure 12 the conversion, X, is plotted as a function of the rate ratio, δ, for different amounts of external heat input. A problem with using membrane reactors for endothermic reactions is that there is heat loss due to permeation. As a result, with decrease in heat input the extent of shift reduces. There is even an optimum rate ratio beyond which the extent of shift is negated by heat loss by permeation.

CONCLUSIONS

1. For a membrane reactor without recycle or backpermeation there is a maximum permeation rate (i.e. minimum thickness of membrane) beyond which the conversion decreases. The maximum conversion attainable is dependent mainly on the permeability of reactants.

2. Cocurrent operation is preferred for cyclohexane dehydrogenation, while countercurrent is more advantageous for propylene disproportionation. For the hydrogen iodide decomposition however, the choice of flow type depends on operating conditions.

3. Equilibrium shift beyond the maximum for the cyclohexane reaction can be achieved by backpermeation of reactant to the catalyst side. For the other two reactions, recycle of permeate is necessary to enhance conversion. For the propylene reaction further improvement in conversion can be achieved by shifting feed location in a manner similar to the Continuous Membrane Column.

4. Temperature effects are mild for the hydrogen iodide and propylene reactions and conversion decrease is not very significant. For the cyclohexane reaction however, due to its high endothermicity, considerable heat has to be supplied in order to achieve reasonable conversion. There is also an optimum permeation rate beyond which equilibrium shift is negated by loss of heat to the permeation side.

SYMBOLS

k^* = pre-exponential constant in forward reaction rate expression

C_P = average specific heat of gas mixture

Da = Damkohler number, kP_hL/F_{A0}

e_T = $\exp[E_A(1/T_0-1/T)/R]$

e_P = $[1.0/(T_r+1)]^{0.5}$

E_A = activation energy for forward reaction, cal/mole

f_r = dimensionless rate expression for reaction

F_i = dimensionless flow of gas on reaction side, (gas flow rate / feed flow rate)

g_i = dimensionless permeation flow rate

h_{ji} = differential material balance ($j=1$ reaction side; $j=2$ permeation side)

ΔH = heat of reaction, cal/gmole

k = forward rate constant, $k^*\exp(-E_A/RT_0)$, mol/m^3.s.Pa

K_e = equilibrium constant

L = length of membrane reactor, m

PE = permeability of fast gas, mol/m. Pa. s (at T_0)

P_h = reaction side pressure, Pa

P_l = permeation side pressure, Pa

Pr = pressure ratio, P_h/P_l

Q_i = dimensionless flow of gas on permeation side, (molar flowrate of gas/ molar feed flow rate)

R = universal gas constant, 1.987 cal/(gmole)(K)

S_m = area of membrane (inner wall) per unit length of membrane reactor, m^2/m

S_w = area of outer wall per unit length of membrane reactor, m^2/m

T = temperature, K

T_a = dimensionless ambient temperature, $(T_{ambient}-T_0)/T_0$

T_r = dimensionless reaction side temperature, $(T-T_0)/T_0$

T_p = dimensionless permeation side temperature, $(T-T_0)/T_0$

t_m = thickness of membrane, m

U_r = overall heat transfer coefficient (reaction side to permeation side) cal/m^2.K. s

U_p = overall heat transfer coefficient (permeation side to outside) cal/m^2.K. s

V_R = volume ratio, reaction side volume of permreactor/volume of impermeable wall plug flow reactor

x_i = mole fraction of gas on the reaction side
X = conversion
X_R = conversion ratio, permreactor conversion/equilibrium conversion
X_{RO} = conversion ratio obtained when $f_r = 0.0$
X_{RM} = limiting conversion ratio attainable in a membrane reactorwhen there is no backpermeation or recycle
ΔX_R = difference in conversion ratio between cocurrent and countercurrent operation
y_i = mole fraction of gas on the permeation side
Z = axial position/length of permreactor

Subscripts

A = reactant
B = slow product
C = fast product
i = reactant, product or inert
0 = permreactor inlet; feed conditions
opt = optimum

Greek letters

α_i = permselectivity of gas
β = dimensionless measure of heat transfer through membrane
= $U_r S_m L/(F_{A0} C_P)$
δ = ratio of permeation rate to reaction rate,$(PE)S_m P_h L/t_m/Da$
ϕ = heat generation,$\Delta H/C_P/T_0$
ψ = dimensionless measure of heat transfer to the environment
= $U_P S_w L/(F_{A0} C_P)$

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