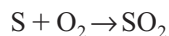


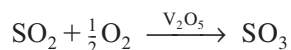
R8.4 Industrial Example of Nonadiabatic Reactor Operation: Oxidation of Sulfur Dioxide

R8.4.1 Manufacture of Sulfuric Acid

In the manufacture of sulfuric acid from sulfur, the first step is the burning of sulfur in a furnace to form sulfur dioxide:



Following this step, the sulfur dioxide is converted to sulfur trioxide, using a catalyst:



A flowsheet of a typical sulfuric acid manufacturing plant is shown in Figure A8-1. It is the converter that we shall be treating in this section.

Although platinum catalysts once were used in the manufacture of sulfuric acid, the only catalysts presently in use employ supported vanadia.¹² For our problem, we shall use a catalyst studied by Eklund, whose work was echoed extensively by Donovan¹³ in his description of the kinetics of SO₂ oxidation. The catalyst studied by Eklund was a Reymersholm V₂O₅ catalyst deposited on a pumice carrier. The cylindrical pellets had a diameter of 8 mm and a length of 8 mm, with a bulk density of 33.8 lb/ft³. Between 818 and 1029°F, the rate law for SO₂ oxidation over this particular catalyst was

$$-r'_{\text{SO}_2} = k \sqrt{\frac{P_{\text{SO}_2}}{P_{\text{SO}_3}}} \left[P_{\text{O}_2} - \left(\frac{P_{\text{SO}_3}}{K_p P_{\text{SO}_2}} \right)^2 \right] \quad (\text{R8.4-1})$$

in which P_i was the partial pressure of species i . This equation can be used when the conversion is greater than 5%. At all conversions below 5%, the rate is essentially that for 5% conversion.

Sulfuric acid manufacturing processes use different types of reactors. Perhaps the most common type has the reactor divided into adiabatic sections with cooling between the sections (recall Figure 8-8). One such layout is shown in Figure R8.4-2. In the process in Figure R8.4-2, gas is brought out of the converter to cool it between stages, using the hot converter reaction mixture to preheat boiler feedwater, produce steam, superheat steam, and reheat the cold gas, all to increase the energy efficiency of the process. Another type has cooling tubes embedded in the reacting mixture. The one illustrated in Figure A8-3 uses incoming gas to cool the reacting mixture.

A typical sulfuric acid plant built in the 1970s produces 1000 to 2400 tons of acid/day.¹⁴ Using the numbers of Kastens and Hutchinson,¹⁵ a 1000-ton/day

An SO₂ flow rate of 0.241 mol/s over 132.158 lb of catalyst can produce 1000 tons of acid per day.

¹²G. M. Cameron, *Chem. Eng. Prog.*, 78(2), 71 (1982).

¹³R. B. Eklund, Dissertation, Royal Institute of Technology, Stockholm, 1956, as quoted by J. R. Donovan, in *The Manufacture of Sulfuric Acid*, ACS Monograph Series 144, W. W. Duecker and J. R. West, eds. (New York: Reinhold, 1959), pp. 166–168.

¹⁴L. F. Friedman, *Chem. Eng. Prog.*, 78(2), 51 (1982).

¹⁵M. L. Kastens and J. C. Hutchinson, *Ind. Eng. Chem.*, 40, 1340 (1948).

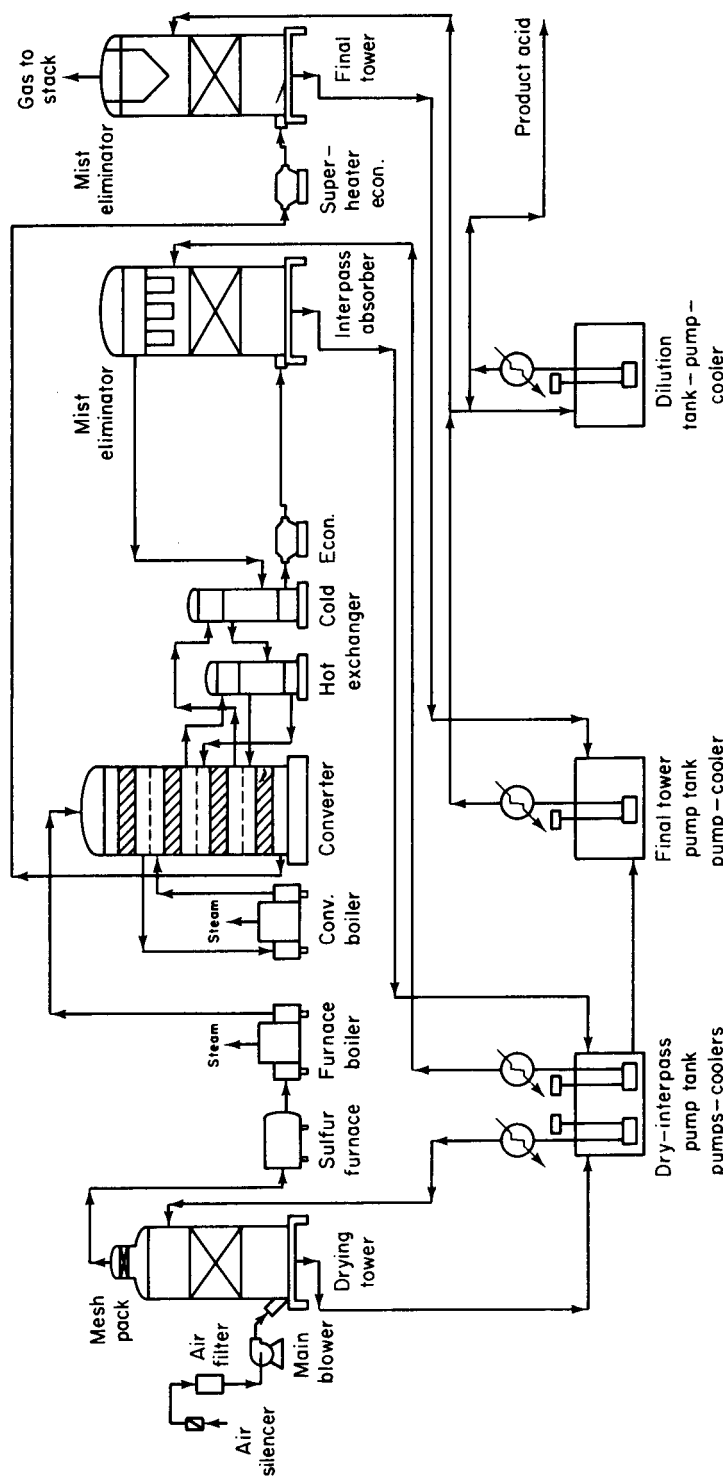


Figure R8.4-1 Flowsheet of a sulfuric acid manufacturing process. [Reprinted with permission of the AIChE and L. J. Friedman. Copyright © 1982 AIChE. All rights reserved.]

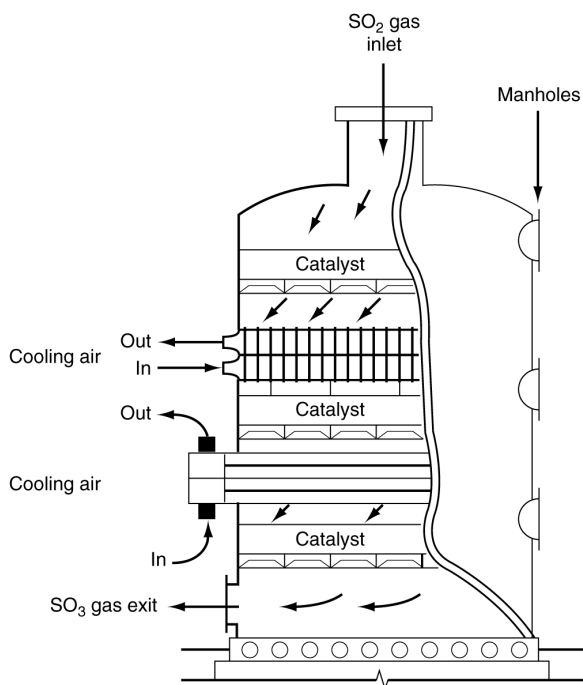


Figure R8.4-2 Sulfur dioxide converter with internal cooling between catalyst layers. [Reprinted with permission of Barnes & Noble Books.]

sulfuric acid plant might have a feed to the SO_2 converter of 7900 lb mol/h, consisting of 11% SO_2 , 10% O_2 , and 79% inerts (principally N_2). We shall use these values.

For preliminary design purposes, we shall calculate the conversions for two situations and compare the results. Only one of the situations will be presented in detail in this example.

1. The first situation concerns two stages of a typical commercial adiabatic reactor. The principles of calculating the conversion in an adiabatic reactor were covered earlier and illustrated in Section 8.3, so will not be presented here but as a problem at the end of the chapter.
2. The second case concerns a reactor with the catalyst in tubes, with the walls cooled by a constant-temperature boiling liquid. Calculations for this system are presented in detail next.

R8.4.2 Catalyst Quantities

Harrer¹⁶ states that the volumetric flow rate in an adiabatic SO_2 converter, measured at normal temperature and pressure, customarily is about 75 to 100

¹⁶T. S. Harrer, in *Kirk-Othmer Encyclopedia of Chemical Technology*, 2nd ed., Vol. 19 (New York: Wiley-Interscience, 1969), p. 470.

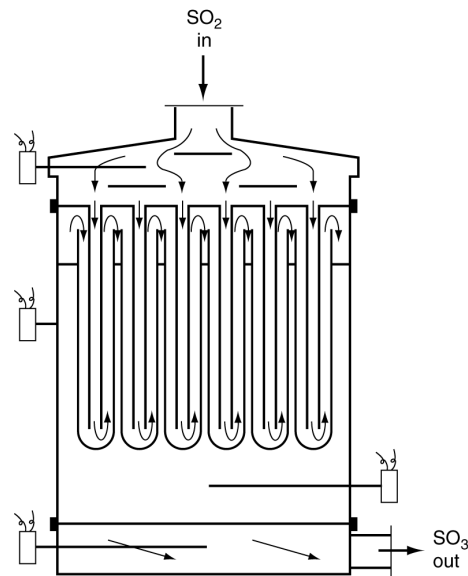


Figure R8.4-3 Sulfur dioxide converter with catalyst cooled by incoming reaction mixture. [Reprinted with permission of Barnes & Noble Books.]

$\text{ft}^3/\text{min} \cdot \text{ft}^2$ of converter area. He also states that the catalyst beds in the converter may be from 20 to 50 in. deep.

It is desirable to have a low mass velocity through the bed to minimize blower energy requirements, so the $75 \text{ ft}^3/\text{min} \cdot \text{ft}^2$ value will be used. Normal conversions in adiabatic converters are 70% in the first stage and an additional 18% in the second.¹⁷ Using Eklund's Reymersholm catalyst, solution of the adiabatic reactor problem at the end of the chapter shows that these conversions require 1550 ft^3 (23 in. deep) in the first stage and 2360 ft^3 (35 in. deep) in the second. As a result, in our cooled tubular reactor, we shall use a total catalyst volume of 3910 ft^3 .

R8.4.3 Reactor Configuration

The catalyst is packed in tubes, and the tubes are put in heat exchangers where they will be cooled by a boiling liquid. The outside diameter of the tubes will be 3 in. Severe radial temperature gradients have been observed in SO_2 oxidation systems,¹⁸ although these systems had platinum catalysts and greatly different operating conditions than those being considered here. The 3-in. diameter is chosen as a compromise between minimizing temperature gradi-

¹⁷J. R. Donovan and J. M. Salamone, in *Kirk-Othmer Encyclopedia of Chemical Technology*, 3rd ed., Vol. 22 (New York: Wiley-Interscience, 1978), p. 190.

¹⁸For example, R. W. Olson, R. W. Schuler, and J. M. Smith, *Chem Eng. Prog.*, 46, 614 (1950); and R. W. Schuler, V. P. Stallings, and J. M. Smith, *Chem. Eng. Prog. Symp. Ser.* 48(4), 19 (1952).

ents and keeping the number of tubes low. For this service, 12-gauge thickness is specified, which means a thickness of 0.109 in. and an inside diameter of 2.782 in. A 20-ft length will be used, as a compromise between decreasing blower energy requirements (shorter tube length) and lowering capital costs (fewer tubes from a longer tube length). For 3910 ft³ of catalyst, the number of tubes that will be required is

$$N_t = \frac{\text{volume of catalyst}}{\text{volume per tube}} = \frac{3910}{(20)(\pi)(2.782/12)^2/4} = 4631 \text{ tubes}$$

The total cross-sectional area of the tubes is

$$A_c = \frac{3910 \text{ ft}^3}{20 \text{ ft}} = 195.5 \text{ ft}^2$$

The overall heat-transfer coefficient between the reacting gaseous mixture and the boiling coolant is assumed to be 10 Btu/h · ft² · °F. This coefficient is toward the upper end of the range of heat-transfer coefficients for such situations as reported by Colburn and Bergelin.¹⁹

R8.4.4 Operating Conditions

Sulfur dioxide converters operate at pressures only slightly higher than atmospheric. An absolute pressure of 2 atm will be used in our designs. The inlet temperature to the reactor will be adjusted so as to give the maximum conversion. Two constraints are present here. The reaction rate over V₂O₅ catalyst is negligible below ~750°F, and the reactor temperature should not exceed ~1125°F at any point.²⁰ A series of inlet temperatures should be tested, and the one above 760°F giving the maximum conversion, yet having no reactor temperature exceeding 1120°F, should be used.

The cooling substance should operate at a high temperature so as to improve thermal efficiency by reuse of heat. The most suitable substance appears to be Dowtherm A, with a normal operating limit of ~750°F but which on occasion has been used as the coolant in this preliminary design.²¹

Example R8.4-1 Oxidation of SO₂

The feed to an SO₂ converter is 7900 lb mol/h and consists of 11% SO₂, 10% O₂, and 79% inerts (principally N₂). The converter consists of 4631 tubes packed with catalyst, each 20 ft long. The tubes are 3 in. o.d. and 2.782 in. i.d. The tubes will be

¹⁹Colburn and Bergelin, in *Chemical Engineers' Handbook*; 3rd ed. (New York: McGraw-Hill, 1950).

²⁰J. R. Donovan and J. M. Salamone, in *Kirk-Othmer Encyclopedia of Chemical Technology*, 3rd ed. (New York: Wiley, 1984).

²¹The vapor pressure of Dowtherm A at 805°F is very high, and this pressure would have to be maintained in the shell side of the reactor for boiling Dowtherm A to be used as a coolant at this temperature. This aspect will be included in the discussion of the problem results.

cooled by a boiling liquid at 805°F, so the coolant temperature is constant over this value. The entering pressure is 2 atm.

For inlet temperatures of 740 and 940°F, plot the conversion, temperature, equilibrium conversion, and reaction rate profile down the reactor.

Additional information:

$$\begin{array}{ll} \phi = 0.45 & U = 10 \text{ Btu/h} \cdot \text{ft}^2 \cdot \text{R} \\ \rho_0 = 0.054 \text{ lb/ft}^3 & A_c = 0.0422 \text{ ft}^2 \\ P_0 = 2 \text{ atm} & T_0 = 1400^\circ\text{R} \text{ (also } T_0 = 1200 \text{ R)} \\ D_p = 0.015 \text{ ft} & g_c = 4.17 \times 10^8 \text{ lb}_m \cdot \text{ft/lb}_f \cdot \text{h}^2 \\ \mu = 0.090 \text{ lb/ft} \cdot \text{h at } 1400 \text{ R} & \rho_b = 33.8 \text{ lb/ft}^3 \text{ (bulk density)} \end{array}$$

Using recent JANAF²² values of K_p at 700 and 900 K, the equilibrium constant at any temperature T is

$$K_p = \exp\left(\frac{42,311}{RT} - 11.24\right) \quad (K_p \text{ in atm}^{-1/2}, T \text{ in R}) \quad (\text{ER8.4-1})$$

at 1600°R,

$$K_p = 7.8 \text{ atm}^{-1/2}$$

For rate constants, the data of Eklund²³ can be correlated very well by the equation

$$k = \exp\left[\frac{-176,008}{T} - (110.1 \ln T) + 912.8\right] \quad (\text{ER8.4-2})$$

where k is in lb mol SO₂/lb cat. · s · atm and T is in R.

There are diffusional effects present in this catalyst at these temperatures, and Equation (E8-10.2) should be regarded as an empirical equation that predicts the effective reaction rate constant over the range of temperatures listed by Donovan (814 to 1138°F). The JANAF tables were used to give the following:

$$\Delta H_{R_x}(800^\circ\text{F}) = -42,471 \text{ Btu/lb mol SO}_2$$

$$C_{p_{\text{SO}_2}} = 7.208 + 5.633 \times 10^{-3} T - 1.343 \times 10^{-6} T^2$$

$$C_{p_{\text{O}_2}} = 5.731 + 2.323 \times 10^{-3} T - 4.886 \times 10^{-7} T^2$$

$$C_{p_{\text{SO}_3}} = 8.511 + 9.517 \times 10^{-3} T - 2.325 \times 10^{-6} T^2$$

$$C_{p_{\text{N}_2}} = 6.248 + 8.778 \times 10^{-4} T - 2.13 \times 10^{-8} T^2$$

where C_p is in Btu/lb mol · °R and T in °R.

Kinetic and
thermodynamic
properties

²²D. R. Stull and H. Prophet, Project Directors, *JANAF Thermochemical Tables*, 2nd ed., NSRDS-NBS 37 (Washington, D.C.: U.S. Government Printing Office, 1971).

²³R. B. Eklund, as quoted by J. R. Donovan, in W. W. Duecker and J. R. West, *The Manufacture of Sulfuric Acid* (New York: Reinhold, 1959).

Solution

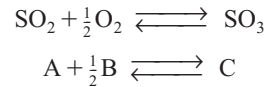
 1. **General procedure:**

- Apply the *plug-flow design equation* relating catalyst weight to the rate of reaction and conversion. Use stoichiometric relationships and feed specifications to express the rate law as a function of conversion.
 - Apply the *energy balance* relating catalyst weight and temperature.
 - Using the *Ergun equation* for pressure drop, determine the pressure as a function of catalyst weight.
 - State *property values* [e.g., k , K_p , $\Delta H_{RX}^\circ(T_R)$, C_{p_i}] and their respective temperature dependences necessary to carry out the calculations.
 - Numerically integrate* the design equation, energy balance, and Ergun equation simultaneously to determine the exit conversion and the temperature and concentration profiles.
2. **Design equations.** The general mole balance equations (design equations) based on the weight of catalyst were given in their differential and integral forms by

$$F_{A0} \frac{dX}{dW} = -r'_A$$

 3. **Rate law:**

$$-r'_{SO_2} = k \sqrt{\frac{P_{SO_2}}{P_{SO_3}}} \left[P_{O_2} - \left(\frac{P_{SO_3}}{K_p P_{SO_2}} \right)^2 \right]$$

 4. **Stoichiometric relationships and expressing $-r'_{SO_2}$ as a function of X :**


We let A represent SO_2 and v_i be the stoichiometric coefficient for species i :

$$P_i = C_i(RT) = C_{A0} \frac{(\Theta_i + v_i X)(RT)P}{(1 + \epsilon X)(T/T_0)P_0} = P_{A0} \frac{(\Theta_i + v_i X)P}{(1 + \epsilon X)P_0} \quad (\text{ER8.4-3})$$

Substituting for partial pressures in the rate law and combining yields

$$\frac{dX}{dW} = \frac{-r'_A}{F_{A0}} = \frac{k}{F_{A0}} \sqrt{\frac{1-X}{\Theta_{SO_3} + X}} \left[\frac{P}{P_0} P_{A0} \frac{\Theta_{O_2} - \frac{1}{2}X}{1 + \epsilon X} - \left(\frac{\Theta_{SO_3} - X}{1 + X} \right)^2 \frac{1}{K_p^2} \right] \quad (\text{ER8.4-4})$$

where $\epsilon = -0.055$, $P_{A0} = 0.22$ atm, $\Theta_{SO_2} = 1.0$, $\Theta_{O_2} = 0.91$, $\Theta_{SO_3} = 0.0$, and $\Theta_{N_2} = 7.17$; $F_{T0} = 7900$ lb mol/h, and $F_{A0} = 869$ lb mol/h.

Per tube:

Weight of catalyst in one tube = $W = \rho_b \pi D^2 / 4L = 28.54$ lb cat./tube

$$F_{A0} = \frac{869}{4631} = 0.188 \text{ lb mol/h} \cdot \text{tube}$$

The combined mole balance, rate law, and stoichiometry

Substituting these values gives us

$$\frac{dX}{dW} = \frac{-r'_A}{F_{A0}} = 5.32k \sqrt{\frac{1-X}{X}} \left\{ \left(\frac{0.2-0.11X}{1-0.055X} \right) \frac{P}{P_0} - \left[\frac{X}{(1-X)K_p} \right]^2 \right\} \quad (\text{ER8.4-5})$$

that is,

$$\frac{dX}{dW} = f_1(X, T, P) \quad (\text{ER8.4-6})$$

The limits of integration are from zero to the weight of catalyst in one tube, 28.54 lb.

5. **Energy balance.** For steady-state operation and no shaft work, Equation (8-56) can be rewritten in terms of catalyst weight as the spatial variable, that is,

$$\frac{dT}{dW} = \frac{(4U/\rho_b D)(T_a - T) + (-r'_A)[- \Delta H_{R_x}(T)]}{F_{A0}(\sum \Theta_i C_{P_i} + X \Delta C_p)} \quad (\text{ER8.4-7})$$

6. **Evaluating the energy balance parameters:**

Heat of reaction:

$$\Delta H_{R_x}(T) = \Delta H_{R_x}^\circ(T_R) + \Delta\alpha(T - T_R) + \frac{\Delta\beta}{2}(T^2 - T_R^2) + \frac{\Delta\gamma}{3}(T^3 - T_R^3) \quad (\text{ER8.4-8})$$

For the SO₂ oxidation, SO₂ + ½O₂ → SO₃,

$$\Delta\alpha = \alpha_{\text{SO}_3} - \frac{1}{2}\alpha_{\text{O}_2} - \alpha_{\text{SO}_2} = 8.511 - (0.5)(5.731) - 7.208 = -1.563$$

Similarly,

$$\Delta\beta = 0.00262 \quad \text{and} \quad \Delta\gamma = -0.738 \times 10^{-6}$$

Substituting into Equation (ER8.4-8) with T_R = 1260°R, we have

$$\Delta H_{R_x}(T) = -42,471 - (1.563)(T - 1260) + (1.36 \times 10^{-3})(T^2 - 1260^2) - (2.459 \times 10^{-7})(T^3 - 1260^3)$$

$$\sum \Theta_i C_{P_i} = 57.23 + 0.014T - 1.94 \times 10^{-6}T^2$$

Heat-transfer coefficient term:

$$\begin{aligned} \frac{U\pi D}{\rho_b A_c} &= \frac{4U}{\rho_b D} = \frac{4(10 \text{ Btu/h} \cdot \text{ft}^2 \cdot ^\circ\text{R})}{(33.8 \text{ lb/ft}^3)(2.78/12) \text{ ft}} \\ &= 5.11 \text{ Btu/h} \cdot \text{lb cat} \cdot \text{R} \end{aligned}$$

Energy balance

$$\frac{dT}{dW} = \frac{5.11(T_a - T) + (-r'_A)[- \Delta H_{R_x}(T)]}{0.188(\sum \Theta_i C_{P_i} + X \Delta C_p)} \quad (\text{ER8.4-9})$$

that is,

$$\frac{dT}{dW} = f_2(T, P, X) \quad (\text{ER8.4-10})$$

7. **Pressure drop:** After rearranging Equation (4-23), the pressure drop is given by

Momentum balance

$$\frac{dP}{dz} = -\frac{(1-\phi)G(1+\varepsilon X)}{\rho_0(P/P_0)(T_0/T)g_c D_p \phi^3} \left[\frac{150\mu(1-\phi)}{D_p} + 1.75G \right]$$

where

$$G = \frac{\sum F_{i0} M_i}{A_c} \quad (M_i = \text{molecular weight of } i)$$

$$= 1307.6 \text{ lb/ft}^2 \cdot \text{h}$$

$$A_c = \text{cross-sectional area } \pi D^2/4$$

Recalling that $W = \rho_b A_c z$, we obtain

$$\frac{dP}{dW} = -\frac{GTP_0(1-\phi)(1+\varepsilon X)}{\rho_b A_c \rho_0 T_0 P D_p g_c \phi^3} \left[\frac{150(1-\phi)\mu}{D_p} + 1.75G \right] \quad (\text{ER8.4-11})$$

8. **Evaluating the pressure-drop parameters:**

$$f_3(X, T, P) = -\frac{GTP_0(1-\phi)(1+\varepsilon X)}{\rho_b A_c \rho_0 T_0 \phi^3 D_p g_c P} \left[\frac{150(1-\phi)\mu}{D_p} + 1.75G \right]$$

Substituting in Equation (ER8.4-11), we get

$$\frac{dP}{dW} = \frac{-1.12 \times 10^{-8}(1-0.55X)T}{P} (5500\mu + 2288) \quad (\text{ER8.4-12})$$

We wish to obtain an order-of-magnitude estimate of the pressure drop. To obtain this estimate, we consider the reaction to be carried out isothermally with $\epsilon = 0$,

$$\boxed{\frac{dP}{dW} = \frac{-0.0432}{P}}$$

Back-of-the-envelope calculation for ΔP

Integrating with limits $P_0 = 2$ atm at $W = 0$ and $P = P$ at $W = 28.54$ lb of catalyst yields

$$\frac{P^2 - 4}{2} = -0.0432(0 - 28.54)$$

$$P = 1.239 \text{ atm}$$

$$\Delta P = 2 - 1.24 = 0.76 \text{ atm}$$

Because the gas-phase viscosity is a weakly varying function of temperature (i.e., $\mu \sim \sqrt{T}$), we shall consider viscosity to be independent of temperature:

$$\frac{dP}{dW} = f_3(T, P, X) \quad (\text{ER8.4-13})$$

9. **Solution procedure.** There are three coupled differential equations that must be solved simultaneously:

$$\text{Mole balance:} \quad \frac{dX}{dW} = f_1(T, P, X) \quad (\text{ER8.4-14})$$

$$\text{Energy balance:} \quad \frac{dT}{dW} = f_2(T, P, X) \quad (\text{ER8.4-15})$$

$$\text{Momentum balance:} \quad \frac{dP}{dW} = f_3(T, P, X) \quad (\text{ER8.4-16})$$

The coupled differential equations to be solved with an ODE solver

10. **Numerical procedure.** The rate equation is independent of conversion between $X = 0.0$ and $X = 0.05$, and the rate of disappearance of SO_2 over this range is equal to the rate of reaction at $X = 0.05$:

$$-r'_{\text{SO}_2} = k \left(0.848 - \frac{0.012}{K_p^2} \right) \quad (\text{ER8.4-17})$$

- Set $X = 0.00$, $T = T_0$, and $P = P_0$.
- Calculate k from Equation (ER8.4-2).
- Calculate K_p from Equation (ER8.4-1).
- If $X < 0.05$, calculate $-r'_{\text{SO}_2}$ from Equation (ER8.4-17). If $X > 0.05$, use Equation (ER8.4-5).
- Calculate X , T , and P from a numerical solution to Equations (ER8.4-5), (ER8.4-9) and (ER8.4-12).

The Polymath program is given in Table ER8.4-1.

11. **Discussion of results.** Figures ER8.4-1.1(a) and (b) show the profiles for inlet temperatures of 1200°R and 1400°R , respectively. Only 68.5% conversion is achieved for $T_0 = 1200^\circ\text{R}$, even though $X_e = 0.99$. For an entering temperature of 1400°R , the major portion of the reaction takes place in the first 6 ft of the reactor. At this point, the conversion is 0.81, with only another 0.06 of the conversion occurring in the remaining 14 ft, as shown in Figure ER8.4-1.1(b). The cause of this low amount of conversion in the final 14 ft is the steadily dropping temperature in the reactor. Beyond the 6-ft point, the temperature is too low for much reaction to take place, which means that the reactor is cooled too much.

This detrimental situation indicates that the coolant temperature is too low for obtaining maximum conversion. Thus even boiling Dowtherm A at its highest possible operating temperature is not a suitable coolant. Perhaps a gas would give a better performance as a coolant in this reaction system. Two problems at the end of the chapter pursue this aspect. One of them seeks the optimum coolant temperature for a constant-coolant-temperature system, and the other uses inlet gas as a coolant.

TABLE ER84-1 SO₃ OXIDATION POLYMATH PROGRAM

Equations:

Variable	initial value	minimal value	maximal value	final value
w	0	0	28.54	28.54
x	0	0	0.9066122	0.9066122
P	2	1.3213547	2	1.3213547
T	1400	1266.8648	1550.2312	1266.8648
fao	0.188	0.188	0.188	0.188
visc	0.09	0.09	0.09	0.09
Ta	1264.67	1264.67	1264.67	1264.67
deltah	-4.237E+04	-4.247E+04	-4.224E+04	-4.247E+04
sum	73.0276	71.852512	74.270983	71.852512
dcp	0.79902	0.6989218	0.8805542	0.6989218
k	0.0981989	0.0107727	0.256642	0.0107727
thetaso	0	0	0	0
Po	2	2	2	2
Pao	0.22	0.22	0.22	0.22
thetao	0.91	0.91	0.91	0.91
eps	-0.055	-0.055	-0.055	-0.055
R	1.987	1.987	1.987	1.987
Kp	52.981141	12.133524	262.00198	262.00198
ra	-0.0832722	-0.1022907	-2.368E-04	-2.368E-04



ODE Report (RK45)

Differential equations as entered by the user

- [1] $d(x)/d(w) = -(ra)/fao$
- [2] $d(P)/d(w) = (-1.12 \cdot 10^{-8}) \cdot (1 - 0.055 \cdot x) \cdot T \cdot (5500 \cdot visc + 2288) / P$
- [3] $d(T)/d(w) = (5.11 \cdot (Ta - T) + (-ra) \cdot (-deltah)) / (fao \cdot (sum + x \cdot dcp))$

Explicit equations as entered by the user

- [1] $fao = .188$
- [2] $visc = .090$
- [3] $Ta = 1264.67$
- [4] $deltah = -42471 - 1.563 \cdot (T - 1260) + .00136 \cdot (T^2 - 1260^2) - 2.459 \cdot 10^{-7} \cdot (T^3 - 1260^3)$
- [5] $sum = 57.23 + .014 \cdot T - 1.94 \cdot 10^{-6} \cdot T^2$
- [6] $dcp = -1.5625 + 2.72 \cdot 10^{-3} \cdot T - 7.38 \cdot 10^{-7} \cdot T^2$
- [7] $k = 3600 \cdot \exp(-176008/T - (110.1 \cdot \ln(T)) + 912.8)$

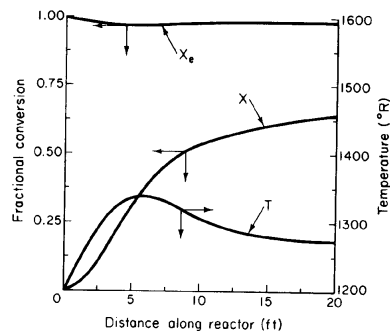


Figure ER8.4-1.1(a) Conversion, temperature, and equilibrium conversion profiles within the reactor: (a) inlet temperature at 1200°R; (b) inlet temperature at 1400°R.

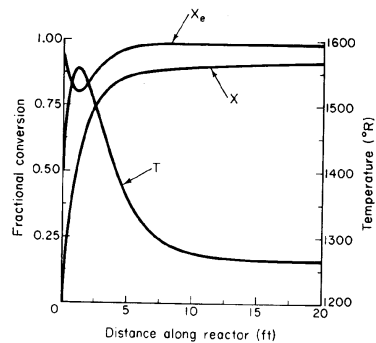


Figure ER8.4-1.1(b) Conversion, temperature, and equilibrium conversion profiles within the reactor: (a) inlet temperature at 1200°R; (b) inlet temperature at 1400°R.

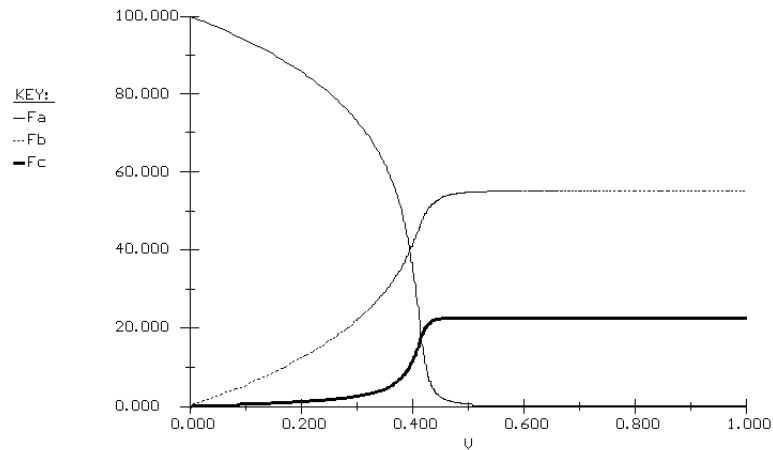


Figure ER84-1.2 Profile of molar flow rates F_A , F_B , and F_C .

Analyzing the
effects of pressure
drop

Another possible way to operate such a reactor is to use multiple-stage operation with progressively higher coolant temperatures. Because pressure drop over the reactor is small (~ 0.7 atm), neglecting the pressure drop does not affect the exit conversion significantly (Figures ER8.4-1.1 or ER8.4-1.2)). The effect is more significant at lower reactor inlet temperatures because the rate of reaction is appreciable over a longer portion of the reactor bed. At higher inlet temperatures, the conversion is limited by the approach to equilibrium, and hence the pressure drop has a negligible effect.

Sec. R8.4 Industrial Example of Nonadiabatic Reactor Operation: Oxidation of Sulfur Diox-