

**CBE 40445**  
**Fall 2020**  
**Test #3**  
**11/4/20**

**1. CSTR vs. Batch reactor  $A \rightarrow M$ , exothermic reaction (60 points)**

Consider a simple first order reaction of  $A \rightarrow M$ . A batch reactor will be compared to a CSTR with the required conversion,  $f_A = .9$  (that is, the exit  $C_A$  concentration/molar flow will be 10% of the feed,  $C_{A0}$ .)

Consider first the batch reactor. Suppose that its volume of  $V = 100$  l is filled with reactant  $A$  that has a concentration,  $C_{A0} = 1$  mole/l. The heat capacity of the  $A$  and  $M$  are the same:  $C_{pA} = C_{pM} = 100$  J/(mole K). The initial temperature is  $T_{\text{feed}} = 300$  K. The reaction rate constant is  $k = 10^{15} \text{Exp}[-12000/T]$  (1/s) and the heat of reaction is  $\Delta H_r = -4,000$  J/mol.

- a. What batch time will be needed for the requisite conversion of 90% of the  $A$  into  $M$  if the temperature is kept at 300K?
- b. What quantity of heat must be removed to keep the reactor at 300K?
- c. If the reactor were operated adiabatically, what would the final temperature be?
- d. For an adiabatic reactor, the maximum reaction **rate** does not always occur at  $f_A = 0$ . Explain why this is so.
- e. Sketch the reactor temperature as a function of time for the adiabatic case.
- f. Under what conditions could the reactor temperature rise and then decrease during the course of a reaction?

Now, independently consider a CSRT with a feed stream of 1 l/s at 300K, where  $C_{A0} = 1$  mol/l. The heat capacity of the  $A$  and  $M$  are the same:  $C_{pA} = C_{pM} = 100$  J/(mole K). The reaction rate constant is  $k = 10^{15} \text{Exp}[-12000/T]$  (1/s) and the heat of reaction is  $\Delta H_r = -4,000$  J/mol.

- g. What reactor volume is needed to accomplish the requisite conversion of 90% of the  $A$  into  $M$  if the temperature is kept at 300K?
- h. If the reactor is operating at 300K, what is the heat removal rate?
- i. If the reactor is operated adiabatically, what is the steady state temperature?
- j. For this case, what is the necessary residence time?

Now compare the batch and CSTR reactors:

- k. How does the temperature rise of the adiabatic batch compare to the CSTR? Will this relation always be the same for other reacting systems?
- l. For an equal amount of  $M$  produced, which reactor requires the greater amount of heat removal to remain isothermal? Explain why.

2. Example 9.3.3 in Davis and Davis — revisited.... (20 points)

Consider accomplishing the reaction  $A + B \Rightarrow C$  in a nonisothermal batch reactor. The reaction occurs in the liquid phase. Find the time necessary to reach 80 percent conversion if the coolant supply is sufficient to maintain the reactor wall at 300 K.

Data:

$$\begin{aligned} C_A^0 &= 0.5 \text{ mol/L} & \Delta H_r &= -15 \text{ kJ/mol} \\ C_B^0 &= 0.6 \text{ mol/L} & UA_H &= 50 \text{ J/(s-K)} \\ C_{p_A} &= C_{p_B} = 65 \text{ J/(mol-K)} & C_{p_C} &= 150 \text{ J/(mol-K)} \\ n_A^0 &= 100 \text{ mol} \\ k &= 5 \times 10^{-3} \exp\left[\frac{20000 \text{ J/mol}}{R_g} \left(\frac{1}{300} - \frac{1}{T}\right)\right] \text{ (L/mol/s)} \end{aligned}$$

■ Answer

The material balance for the batch reactor is:

$$\frac{df_A}{dt} = kC_A^0(1 - f_A)(1.2 - f_A)$$

The energy balance can be written as:

$$UA_H(T^* - T) = \Delta H_r n_A^0 \frac{df_A}{dt} + [n_A^0(1 - f_A)C_{p_A} + n_A^0(1.2 - f_A)C_{p_B} + n_A^0 f_A C_{p_C}] \frac{dT}{dt}$$

The material and energy balance equations must be solved simultaneously. A convenient form for solution by numerical techniques is:

$$\begin{aligned} \frac{df_A}{dt} &= g(T, f_A) = k(T)C_A^0(1 - f_A)(1.2 - f_A) \\ \frac{dT}{dt} &= \frac{UA_H(300 - T) - \Delta H_r n_A^0 g(T, f_A)}{n_A^0(1 - f_A)C_{p_A} + n_A^0(1.2 - f_A)C_{p_B} + n_A^0 f_A C_{p_C}} \end{aligned}$$

with  $f_A = 0$  and  $T = 300$  K at  $t = 0$ . The results are shown in Figure 9.3.3. From the data given in Figure 9.3.3, it takes 462 s to reach a fractional conversion of 0.8. Additionally, the final temperature is 332 K. Notice that the final temperature is not the maximum temperature achieved during the reaction, in contrast to adiabatic operation.

You will recall the homework problem which considers the adiabatic case,  $UA_H=0$  for which the reactor heats up to 389K in 462 s and the conversion increases to 0.90.

However, the last part of the question refers back to this problem and asks for the heat that is removed during the batch reaction.

There are two ways that this might be computed. First is to integrate the heat removal expression,

$$Q = \int_0^t \dot{Q} dt = \int_0^t U A_H (300 - T) dt,$$

when this is done, the result is  $Q = -742$  kJ.

You might also expect to be able to get the heat removed by knowing the heat produced by the reaction and the change in enthalpy of the products and reactants.

$$\text{heat released} = \Delta H f_A n_A^0 = -15000 \text{ kJ/mol} \cdot 100 \text{ mol} \cdot 0.8 = -1.2 \times 10^6 \text{ kJ}$$

Then add the heat soaked up by the final mixture compared to the feed

$$n_A^0 ((1 - f_A) C_{PA} + (1.2 - f_A) C_{PB} + f_A C_{PC}) (332 - 300),$$

The result is -698 kJ.

For this last part we take advantage of enthalpy being a state function and assume the reaction occurs at  $T = 300$ . Then we heat the final products up to 332K.

We see that these answers don't match. A check of the numerics verifies that the  $Q$  from the numerical integral is not significantly in error.

We would like to resolve the difference between these two numbers.

Note that:

1. The *concept* of getting the enthalpy change to measure heat removed is correct.
2. The error persists even if  $k$  is assumed not a function of  $T$ .
3. The error persists even if the initial feed of  $A$  and  $B$  is such that they are both 100 moles.
4. The error goes away if the heat capacity of  $C$  is changed to 130 J/mole-K

- a. Write down the differential equation that gives the change in temperature as a function directly of fractional conversion.
- b. What is the conceptual error with the example and how can it be fixed?

### 3. Plug flow reactor problems problems. (20 points)

Because of available space in the plant, two supposedly identical gas-phase, plug flow reactors were installed as one that is oriented horizontally (i.e., its flow is perpendicular to gravity) and one that is vertical with its flow downward in the same direction as gravity.

Both are 5 meters long and consist of 20, 3 cm diameter tubes filled with a spherical catalyst packing.

The reaction is  $A \rightarrow M$  with a design specification of 99% conversion of  $A$  to  $M$ .

For the first few years of service, (including regular catalyst regeneration with hydrogen) both reactors performed up to spec giving essentially identical conversions.

However after some time, the horizontal reactor began produce substantially (>20%) lower conversion.

- a. What could be going wrong with the horizontal reactor?
- b. What measurement other than inlet and exit concentrations could give some insight into the nature of the problem?

Test 3 additions

Exit time distribution:

$$E(t) = \frac{C(t)}{\int_0^\infty C(\bar{t})d\bar{t}}$$

$$\int_0^\infty E(\bar{t})d\bar{t} = 1$$

5

Energy balance a batch reactor;

$$\int_0^t UA_H(T^* - T)d\bar{t} = \frac{-\Delta H_r|_{T^0}}{\nu_\ell} n_\ell^0 f_\ell + \sum_i \left( n_i \int_{T_{\text{initial}}}^{T_{\text{final}}} C_{p_i} dT \right) \quad (9.3.8)$$

or in differential form:

$$UA_H(T^* - T) = \frac{-\Delta H_r|_{T^0}}{\nu_\ell} n_\ell^0 \frac{df_\ell}{dt} + \sum_i (n_i C_{p_i}) \frac{dT}{dt} \quad (9.3.9)$$

Notice that [see Equations (1.2.10) and (1.3.2)]:

$$\frac{n_\ell^0}{-\nu_\ell} \frac{df_\ell}{dt} = \frac{d\Phi}{dt} = rV \quad (9.3.10)$$

Energy balance for a plug flow reactor:

**Figure 9.4.1 |**  
Schematic of differential fluid volume in a nonisothermal PFR.

where  $d_t$  is the diameter of the tubular reactor. Recall again that the enthalpy contains both sensible heat and heat of reaction effects. Thus, the energy balance Equation (9.2.7) can be written for the differential fluid element of the PFR as:

$$d\dot{Q} = \sum_i \left( F_i^{\text{out}} \int_{T^0}^{T_{\text{out}}} C_{p_i} dT \right) - \sum_i \left( F_i^{\text{in}} \int_{T^0}^{T_{\text{in}}} C_{p_i} dT \right) - \frac{\Delta H_r|_{T^0}}{\nu_\ell} F_\ell^0 df_\ell \quad (9.4.2)$$

where the heat of reaction is evaluated at a reference temperature  $T^0$  and  $C_{p_i}$  is the molar heat capacity of species  $i$ . Normally,  $T^0$  is taken as the reactor entrance temperature.

$$r_\ell V = \frac{F_\ell^0 df_\ell}{-\nu_\ell}$$

Energy balance for a CSTR:  $r_1 V = \frac{F_1^0 f_1^f}{-v_1}, f_1^0 = 0$

energy balance for a CSTR can be derived from Equation (9.2.7) by again  
 out the reaction isothermally at the inlet temperature and then evaluating  
 heat effects at reactor outlet conditions, that is,

$$\dot{Q} = \frac{F_\ell^0 (\Delta H_r|_{T^0})(f_\ell^f - f_\ell^0)}{(-v_\ell)} + \sum \left( F_i^f \int_{T^0}^{T^f} C_{p_i} dT \right) \quad (9.5.1)$$

the superscript  $f$  denotes the final or outlet conditions. For adiabatic opera-  
 : 0.

Test 2 additions to the formula sheet

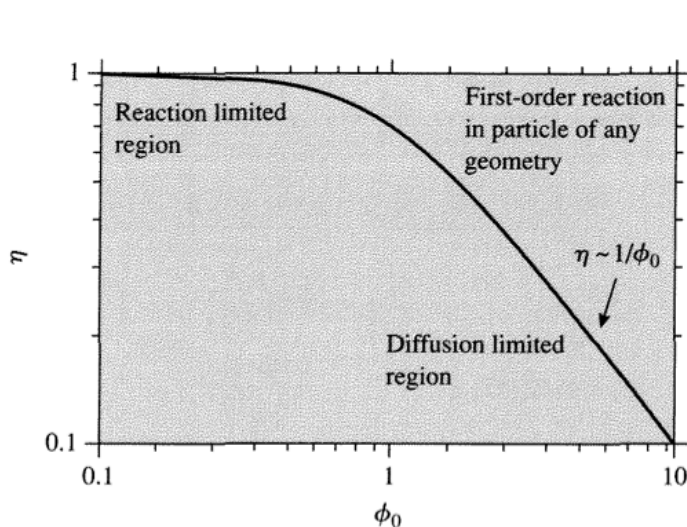
$$N_{Ax} = -D_{AB} \frac{dC_A}{dx} \quad Sh = \frac{\bar{k}_c (2R_p)}{D_{AB}}$$

$$-\frac{dN_A}{dx} = \frac{2k_s}{R_{\text{pore}}} C_A \quad N_A = -D_{TA} \frac{dC_A}{dx}$$

$$\frac{d^2C_A}{dx^2} - \frac{2k_s}{D_{TA}R_{\text{pore}}} C_A = 0 \quad \frac{d^2C_A}{d\chi^2} - \phi^2 C_A = 0$$

$$r_{\text{obs}} = 4\pi (R_p)^2 D_{TA}^e \left. \frac{dC_A}{d\bar{r}} \right|_{\bar{r}=R_p}$$

$$\eta = \frac{r_{\text{obs}}}{r_{\text{max}}} = \frac{4\pi (R_p)^2 D_{TA}^e \left. \frac{dC_A}{d\bar{r}} \right|_{\bar{r}=R_p}}{\frac{4}{3} \pi (R_p)^3 k C_{AS}} = \frac{3}{R_p} \frac{D_{TA}^e}{k} \left. \frac{dC_A}{d\bar{r}} \right|_{\bar{r}=R_p} C_{AS}$$



$$\phi_0 = L_p \sqrt{\frac{k}{D_{TA}^e}}$$

$$L_p = \frac{V_p}{S_p}$$

$$\eta = \frac{\tanh(\phi_0)}{\phi_0}$$

**Figure 6.3.9 |**

Effectiveness factor [ $\eta = \tanh(\phi_0)/\phi_0$ ] for a first-order reaction in a catalyst as a function of the Thiele modulus with generalized length parameter.

**Table 6.3.2** | Characteristic length parameters of common pellet shapes.

	$L_p$
Slab	Length of pore, $x_p$
Cylinder	$R_p/2$
Sphere	$R_p/3$

$$\phi_0 = \frac{V_p}{S_p} \sqrt{\frac{k}{D_{TA}^e}} = \left[ \frac{R_p}{\frac{R_p}{x_p} + 2} \right] \sqrt{\frac{k}{D_{TA}^e}}$$

$$\phi_0 = \frac{V_p}{S_p} \sqrt{\frac{n+1}{2} \frac{kC_{AS}^{n-1}}{D_{TA}^e}} \quad n > -1$$

$$\Gamma_{\text{obs}} = \eta k C_{AS}^n$$



Potentially useful formulae:

$$\Delta G - \Delta G^0 = R_g T \sum_{i=1}^{NCOMP} \nu_i \ln a_i = R_g T \ln \left( \prod_i a_i^{\nu_i} \right) \quad \Delta G^0 = -R_g T \ln \left[ \frac{a_W^w a_S^s \dots}{a_A^a a_B^b \dots} \right] = -R_g T \ln K_a$$

Equilibrium constant:

$$\left[ \frac{\partial (\ln K_a)}{\partial T} \right]_P = \frac{\Delta H^0}{R_g T^2} \quad K_a = \exp \left[ -\Delta G^0 / (R_g T) \right]$$

$$a = \bar{f}_i / \bar{f}_i^0 \quad \bar{f}_i = X_i \bar{\phi}_i P \quad K_a = \left[ \frac{X_W^w X_S^s \dots}{X_A^a X_B^b \dots} \right] \left[ \frac{\bar{\phi}_W^w \bar{\phi}_S^s \dots}{\bar{\phi}_A^a \bar{\phi}_B^b \dots} \right] P^{\bar{w} + \bar{s} + \dots - \bar{a} - \bar{b} \dots}$$

The fractional conversion is defined as:

$$f(t) = \frac{\Phi(t)}{\Phi_{\max}} \quad (1.2.9)$$

and can be calculated from Equations (1.2.3) and (1.2.8):

$$f_\ell = (-\nu_\ell) \frac{\Phi}{n_\ell^0} = 1 - \frac{n_\ell}{n_\ell^0} \quad (1.2.10)$$

Equation (1.2.10) can be rearranged to give:

$$n_\ell = n_\ell^0 (1 - f_\ell) \quad (1.2.11)$$

where  $0 \leq f_\ell \leq 1$ . When the thermodynamics of the system limit  $\Phi$  such that it cannot reach  $\Phi_{\max}$  (where  $n_\ell = 0$ ),  $\Phi$  will approach its equilibrium value  $\Phi^{\text{eq}}$  ( $n_\ell \neq 0$  value of  $n_\ell$  determined by the equilibrium constant). When a reaction is limited by

This expression is either positive or equal to zero if the system has reached equilibrium. The reaction rate, like  $\Phi$ , is an extensive property of the system. A specific rate called the *volumic rate* is obtained by dividing the reaction rate by the volume of the system,  $V$ :

$$r = \frac{1}{V} \frac{d\Phi}{dt} \quad \left( \frac{\text{mol}}{\text{time} \cdot \text{volume}} \right) \quad (1.3.2)$$

Differentiation of Equation (1.2.3) gives:

$$dn_i = \nu_i d\Phi \quad (1.3.3)$$

Substitution of Equation (1.3.3) into Equation (1.3.2) yields:

$$r = \frac{1}{\nu_i V} \frac{dn_i}{dt} \quad (1.3.4)$$

tion (1.2.1), then the number of moles of component  $A_i$  at any time  $t$  will be given by the equation:

$$n_i(t) = n_i^0 + \nu_i \Phi(t) \quad (1.2.3)$$

that is an expression of the *Law of Definitive Proportions* (or more simply, a mole balance) and defines the parameter,  $\Phi$ , called the *extent of reaction*. The extent of reaction is a function of time and is a natural reaction variable.

concentrations,  $C_i$ , is desired. Since  $C_i = n_i/V$ , Equation (1.3.4) can be written as:

$$r = \frac{1}{\nu_i V} \frac{d}{dt} (C_i V) = \frac{1}{\nu_i} \frac{dC_i}{dt} + \frac{C_i}{\nu_i V} \frac{dV}{dt} \quad (1.3.6)$$

Note that only when the volume of the system is constant that the volumic rate can be written as:

$$r = \frac{1}{\nu_i} \frac{dC_i}{dt}, \text{ constant } V \quad (1.3.7)$$

When it is not possible to write a stoichiometric equation for the reaction, the

#### **Rule II**

The rate of an irreversible (one-way) reaction can generally be written in the form:

$$r = k(T) \bar{F}(C_i, T) \quad (1.4.1)$$

where  $\bar{F}(C_i, T)$  is a function that depends on the composition of the system as expressed by the concentrations  $C_i$ .

#### **Rule III**

The rate constant generally depends on the absolute temperature,  $T$ , following the law first proposed by Arrhenius in 1889:

$$k = \bar{A} \exp \left[ \frac{-E}{R_i T} \right] \quad (1.4.2)$$

#### **Rule IV**

Frequently, the function  $\bar{F}(C_i)$  in the expression  $r = k \bar{F}(C_i)$  can approximately be written as:

$$\bar{F}(C_i) = \prod_i C_i^{\alpha_i}$$

Consider the unimolecular reaction:



Using the Guldberg-Waage form of the reaction rate to describe this reaction gives:

$$r = kC_A \quad (1.5.2)$$

From Equations (1.3.4) and (1.5.2):

$$r = \frac{1}{v_i V} \frac{dn_i}{dt} = \frac{-1}{V} \frac{dn_A}{dt} = kC_A$$

$$\frac{dn_A}{dt} = -kn_A \quad (\text{variable } V) \quad (1.5.3)$$

$$\frac{df_A}{dt} = k(1 - f_A) \quad [\text{using Equation (1.2.11)}] \quad (1.5.4)$$

Consider the bimolecular reaction:



Using the Guldberg-Waage form of the reaction rate to describe this reaction gives:

$$r = kC_A C_B \quad (1.5.8)$$

$$r = \frac{1}{v_i V} \frac{dn_i}{dt} = \frac{-1}{V} \frac{dn_A}{dt} = kC_A C_B$$

$$r = \left( \frac{\bar{k}T}{h} \right) \exp\left[ \frac{\Delta S_0^\ddagger}{R_g} \right] \exp\left[ \frac{-\Delta H_0^\ddagger}{R_g T} \right] C_A C_B$$

$$\frac{dn_i}{dt} = 0 - 0 + (v_i r V)$$

accumulation                  input                  output                  amount produced by reaction

The material balance can also be written in terms of the fractional conversion and it is:

$$n_i^0 \frac{df_i}{dt} = -(v_i r) V^0 (1 + \epsilon_i f_i) \quad \text{with } f_i = 0 @ t = 0 \quad (3.2.2)$$

where  $|\epsilon_i| > 0$  for nonconstant volume.

Since the reactor is at steady-state, the difference in  $F_i^0$  (input) and  $F_i$  (output) must be due to the reaction. (In this text, the superscript 0 on flow rates denotes the input to the reactor.) The material balance on a CSTR is written as:

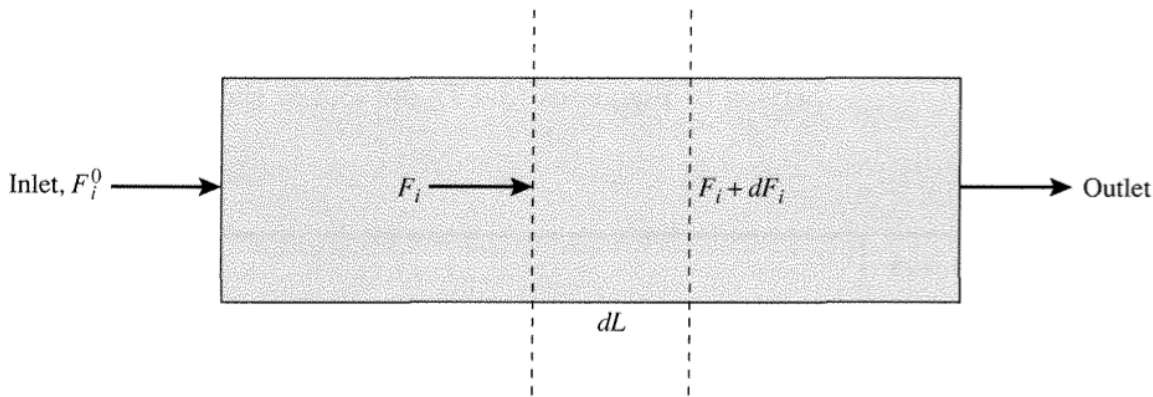
$$0 = F_i^0 - F_i + (v_i r) V \quad (3.3.1)$$

accumulation
input of  $i$ 
output of  $i$ 
amount produced by reaction

$$V \frac{dC_i}{dt} = q_i C_{i0} - q C_i + v_i r V$$

$$f_i = \frac{n_i^0 - n_i}{n_i^0} = \frac{F_i^0 - F_i}{F_i^0}$$

3.2.4) into Equation (3.2.2) yields:



**Figure 3.4.1 |**  
Tubular reactor.

$$0 = F_i - (F_i + dF_i) + v_i r A_C dL \quad (3.4.1)$$

(accumulation)
= (input)
- (output)
+(amount produced by reaction)

where  $A_C$  is the cross-sectional area of the tube. Also,  $A_C dL = dV_R$ , so Equation (3.4.1) can be written as:

$$\frac{dF_i}{dV_R} = v_i r \quad (3.4.2)$$

or

$$F_i^0 \frac{df_i}{dV_R} = -v_i r \quad (3.4.3)$$

Integration of Equation (3.4.3) gives:

$$\frac{V_R}{F_i^0} = \int_{f_i^0}^{f_i^{\text{outlet}}} \frac{df_i}{(v_i r)} \quad \text{or} \quad \tau = \frac{V_R}{v} = C_i^0 \int_{f_i^0}^{f_i^{\text{outlet}}} \frac{df_i}{(v_i r)} \quad (3.4.4)$$

If changes in volume due to reaction are negligible, then  $[F_i = C_i v$ ; moles of  $i$ /time = (moles of  $i$ /volume) (volume/time)]:

$$\frac{dC_i}{d\tau} = \frac{dC_i}{d(V_R/v)} = v_i r \quad (3.4.5)$$

$$x = \frac{-b \pm \sqrt{b^2 - 4ac}}{2a}$$

$$-\int_{C_{A0}}^{C_A} \frac{1}{\xi^2} d\xi = \frac{1}{C_A} - \frac{1}{C_{A0}}$$

$$\int_{x_0}^x \frac{1}{(\xi - \alpha)} d\xi = \ln \left( \frac{x - \alpha}{x_0 - \alpha} \right)$$

$$\int_{x_0}^x \frac{1}{\xi} d\xi = \ln \left( \frac{x}{x_0} \right)$$

## Periodic table of the elements

group	1*	2											13	14	15	16	17	18
period 1	1 H																	2 He
2	3 Li	4 Be											5 B	6 C	7 N	8 O	9 F	10 Ne
3	11 Na	12 Mg	3	4	5	6	7	8	9	10	11	12	13 Al	14 Si	15 P	16 S	17 Cl	18 Ar
4	19 K	20 Ca	21 Sc	22 Ti	23 V	24 Cr	25 Mn	26 Fe	27 Co	28 Ni	29 Cu	30 Zn	31 Ga	32 Ge	33 As	34 Se	35 Br	36 Kr
5	37 Rb	38 Sr	39 Y	40 Zr	41 Nb	42 Mo	43 Tc	44 Ru	45 Rh	46 Pd	47 Ag	48 Cd	49 In	50 Sn	51 Sb	52 Te	53 I	54 Xe
6	55 Cs	56 Ba	57 La	72 Hf	73 Ta	74 W	75 Re	76 Os	77 Ir	78 Pt	79 Au	80 Hg	81 Tl	82 Pb	83 Bi	84 Po	85 At	86 Rn
7	87 Fr	88 Ra	89 Ac	104 Rf	105 Db	106 Sg	107 Bh	108 Hs	109 Mt	110 Ds	111 Rg	112 Cn	113 Nh	114 Fl	115 Mc	116 Lv	117 Ts	118 Og
lanthanoid series 6	58 Ce	59 Pr	60 Nd	61 Pm	62 Sm	63 Eu	64 Gd	65 Tb	66 Dy	67 Ho	68 Er	69 Tm	70 Yb	71 Lu				
actinoid series 7	90 Th	91 Pa	92 U	93 Np	94 Pu	95 Am	96 Cm	97 Bk	98 Cf	99 Es	100 Fm	101 Md	102 No	103 Lr				

\*Numbering system adopted by the International Union of Pure and Applied Chemistry (IUPAC).

© Encyclopædia Britannica, Inc.

**UNITS**

Mass	1 kg = 1000 g = 0.001 metric ton = 2.20462 lb <sub>m</sub> 1 lb <sub>m</sub> = 16 oz = 5E-4 ton = 453.59 g = 0.45359 kg
Length	1 m = 100 cm = 1000 mm = 1E6 μm = 1E9 nm = 39.370 in = 3.2808 ft = 1.0936 yd 1 ft = 12 in = 0.30480 m = 30.480 cm 1 in = 2.5400 cm
Volume	1 m <sup>3</sup> = 1000 L = 1E6 cm <sup>3</sup> = 1E6 ml = 35.315 ft <sup>3</sup> = 264.17 gal 1 ft <sup>3</sup> = 1728.0 in <sup>3</sup> = 7.4805 gal = 0.028317 m <sup>3</sup> = 28.317 L = 28317 ml
Force	1 N = 1 kg·m/s <sup>2</sup> = 1E5 dynes = 1E5 g·cm/s <sup>2</sup> = 0.22481 lb <sub>f</sub> 1 lb <sub>f</sub> = 32.174 lb <sub>m</sub> ·ft/s <sup>2</sup> = 4.4482 N
Pressure	1 atm = 1.01325E5 N/m <sup>2</sup> (Pa) = 1.01325 bar = 760 mmHg at 0°C = 33.9 ftH <sub>2</sub> O at 4°C 1 atm = 14.696 psia 1 bar = 0.1 MPa = 0.98692 atm = 14.504 psia = 750.06 mmHg at 0°C = 10.197 mH <sub>2</sub> O at 4°C
Energy	1 J = 1 N·m = 1 MPa·cm <sup>3</sup> = 1 kgm <sup>2</sup> /s <sup>2</sup> = 0.23901 cal = 0.73756 ft·lb <sub>f</sub> 1 J = 1E7 ergs = 1E7 g·cm <sup>2</sup> /s <sup>2</sup> 1 kJ = 0.94781 Btu <sup>(see note 1)</sup> = 2.7778E-4 kW·h = 0.23901 food calorie
Power	1 W = 1 J/s = 0.2390 cal/s <sup>(see note 2)</sup> = 0.73756 ft·lb <sub>f</sub> /s = 3.4121 Btu/h <sup>(see note 1)</sup> 1 hp = 550 ft·lb <sub>f</sub> /s = 0.70726 Btu/s <sup>(see note 1)</sup> = 0.74570 kW
Gas Constant, <i>R</i>	= 8.31447 J/mole·K = 8.31447 cm <sup>3</sup> ·MPa/mole·K = 8.31447 m <sup>3</sup> ·Pa/mole·K = 8,314.47 cm <sup>3</sup> ·kPa/mole·K = 83.1447 cm <sup>3</sup> ·bar/mole·K = 1.9859 Btu/lbmole·R <sup>(see note 1)</sup> = 82.057 cm <sup>3</sup> ·atm/mole·K = 1.9872 cal/mole·K <sup>(see note 2)</sup> = 10.731 ft <sup>3</sup> ·psia/lbmole·R
Gravitational Constants at sea level	$g = 9.8066 \text{ m/s}^2$ $g/g_c = 9.8066 \text{ N/kg}$ $g_c = 1 \text{ (kg·m/s}^2\text{)/N}$ $g = 32.174 \text{ ft/s}^2$ $g/g_c = 1 \text{ lb}_f\text{/lb}_m$ $g_c = 32.174 \text{ (lb}_m\text{·ft/s}^2\text{)/lb}_f$
Faraday's Constant	$F = 96,485 \text{ J/V}$
IUPAC Standard Conditions of Temperature and Pressure	$T = 0^\circ\text{C} = 273.15 \text{ K}$ ; $P = 0.1 \text{ MPa}$ ; $V(\text{ideal gas}) = 22711 \text{ cm}^3\text{/mole}$ ; $\rho(\text{water}) = 0.99984 \text{ g/cm}^3 = 8.3441 \text{ lb}_m\text{/gal}$

1. The International Steam Table (IT) BTU.
2. The thermochemical calorie.