Chapter 2

Refrigerated Condensers

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2.1   Introduction

Condensers in use today may fall in either of two categories: refrigerated or non-refrigerated. Non-refrigerated condensers are widely used as raw material, product, and/or solvent recovery devices in chemical process industries. They are frequently used prior to control devices (e.g., incinerators or absorbers). Refrigerated condensers are used as air pollution control devices for treating emission streams with high VOC concentrations (usually > 5,000 ppmv) in applications involving gasoline bulk terminals, storage, etc. Condenser control technology is unique in that it not only reduces emissions to the atmosphere but also captures or recovers VOCs, potentially for additional use.

Condensation is a separation technique in which one or more volatile components of a vapor mixture are separated from the remaining vapors through saturation followed by a phase change. The phase change from gas to liquid can be achieved in two ways: (a) the system pressure can be increased at a given temperature, or (b) the temperature may be lowered at a constant pressure. In a two-component system where one of the components is noncondensible (e.g., air), condensation occurs at dew point (saturation) when the partial pressure of the volatile compound is equal to its vapor pressure. The more volatile a compound is (e.g., the lower the normal boiling point), the larger the amount that can remain as vapor at a given temperature; hence the lower the temperature required for saturation (condensation). Refrigeration is often employed to obtain the low temperatures required for acceptable removal efficiencies. This chapter is limited to the evaluation of refrigerated condensation at constant (atmospheric) pressure.

2.1.1   System Efficiencies and Performance

The removal efficiency of a condenser is dependent on the emission stream characteristics including the nature of the VOC in question (vapor pressure/temperature relationship), VOC concentration, and the type of coolant used.[1] Any component of any vapor mixture can be condensed if brought to a low enough temperature and allowed to come to equilibrium. Figure 2.1 shows the vapor pressure dependence on temperature for selected compounds (Erikson, 1980). A condenser cannot lower the inlet concentration to levels below the saturation concentration at the coolant temperature. Removal efficiencies of approximately 50 to 90 percent can be achieved with coolants such as chilled water and brine solutions, and removal efficiencies above 90 percent can be achieved with ammonia, liquid nitrogen, chlorofluorocarbons, hydrochlorofluorocarbons, or hydrofluorocarbons, depending on the VOC composition and concentration level of the emission stream.

2.2   Process Description

Figure 2.2 depicts a typical configuration for a refrigerated surface condenser system as an emission control device. The basic equipment required for a refrigerated condenser system includes a VOC condenser, a refrigeration unit(s), and auxiliary equipment (e.g., precooler, recovery/storage tank, pump/blower, and piping).

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[1] Organic hazardous air pollutants (HAPs) are considered a subset of VOC compounds.
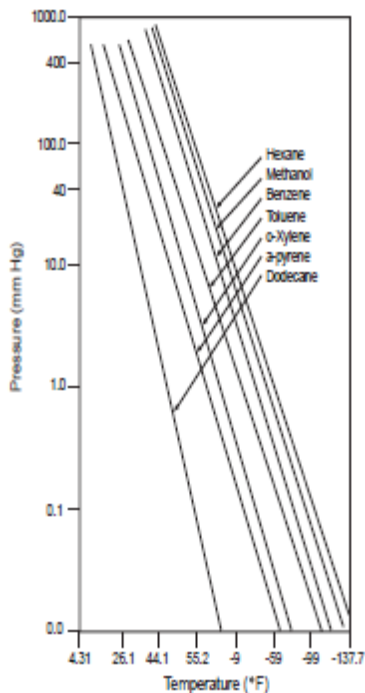
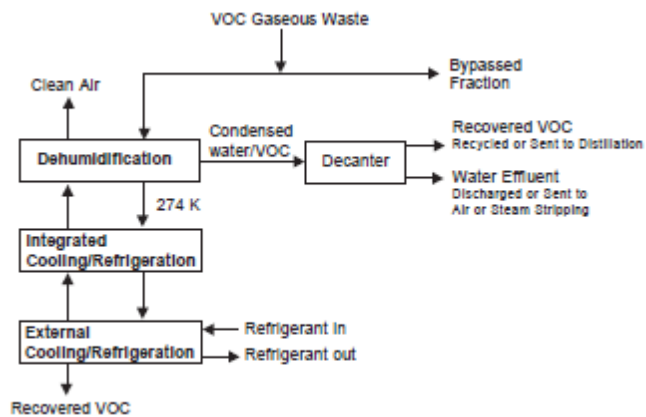**Figure 2.1:** Vapor Pressures of Selected Compounds vs. Temperature (Erikson, 1980)

Figure 2.2: Schematic Diagram for a Refrigerated Condenser System

2.2.1 VOC Condensers

The two most common types of condensers used are surface and contact condensers (Vatavuk and Neveril, 1983). In surface condensers, the coolant does not contact the gas stream. Most surface condensers in refrigerated systems are the shell and tube type (see Figure 2.3) (McCabe and Smith, 1976). Shell and tube condensers circulate the coolant through tubes. The VOCs condense on the outside of the tubes (i.e., within the shell). Plate and frame type heat exchangers are also used as condensers in refrigerated systems. In these condensers, the coolant and the vapor flow separately over thin plates. In either design, the condensed vapor forms a film on the cooled surface and drains away to a collection tank for storage, reuse, or disposal.

In contrast to surface condensers where the coolant does not contact either the vapors or the condensate, contact condensers cool the vapor stream by spraying either a liquid at ambient temperature or a chilled liquid directly into the gas stream.

Spent coolant containing the VOCs from contact condensers usually cannot be reused directly and can be a waste disposal problem. Additionally, VOCs in the spent coolant cannot be directly recovered without further processing. Since the coolant from surface condensers does not contact the vapor stream, it is not contaminated and can be recycled in a closed loop. Surface condensers also allow for direct recovery of VOCs from the gas stream. This chapter addresses the design and costing of refrigerated surface condenser systems only.
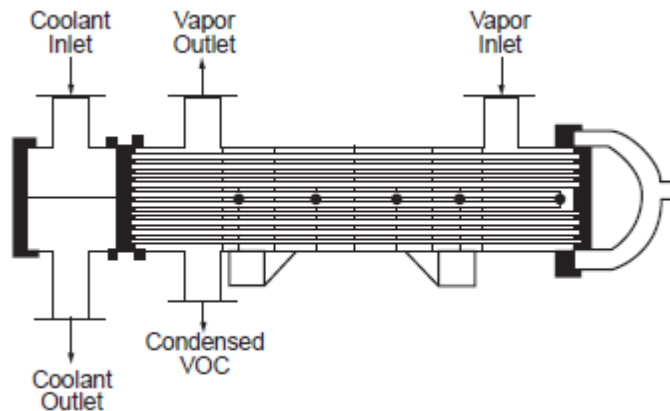

Figure 2.3: Schematic Diagram of a Shell and Tube Surface Condenser (McCabe and Smith, 1976)

2.2.2 Refrigeration Unit

The commonly used mechanical vapor compression cycle to produce refrigeration consists of four stages: evaporation, compression, condensation, and expansion (see Figure 2.4) (Green, 2008). The cycle that is used for single-stage vapor compression involves two pressures, high and low, to enable a continuous process to produce a cooling effect. Heat absorbed from the gas stream evaporates the liquid coolant (refrigerant). Next, the refrigerant (now in vapor phase) is compressed to a higher temperature and pressure by the system compressor. Then, the superheated refrigerant vapor is condensed, rejecting its sensible and latent heat in the condenser. Subsequently, the liquid refrigerant flows from the condenser through the expansion valve, where pressure and temperature are reduced to those in the evaporator, thus completing the cycle.
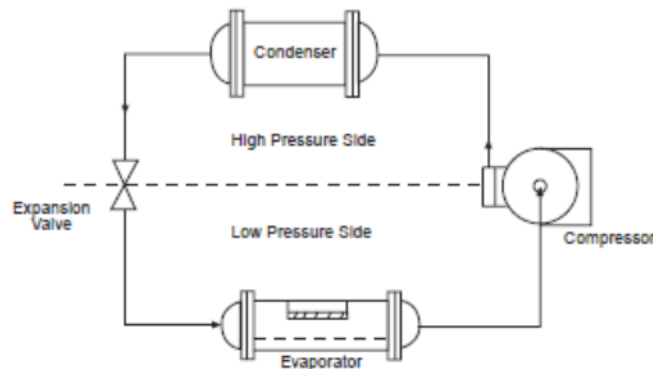
Figure 2.4: Basic Refrigeration Cycle (Green, 2008)

The capacity of a refrigeration unit is the rate at which heat is removed, expressed in tons of refrigeration. One ton of refrigeration is the refrigeration produced by melting one ton of ice at 32°F in 24 hours. It is the rate of removing heat equivalent to 12,000 British thermal units per hour (Btu/hr) or 200 Btu/min. For more details on refrigeration principles, see References Kern, 1950, and Smith, VanNess, and Abbott, 2005.

For applications requiring low temperatures (below about -30°F), multistage refrigeration systems are frequently employed (Green, 2008). Multistage systems are designed and marketed in two different types—compound and cascade. In compound systems, only one refrigerant is used. In a cascade system, two or more separate refrigeration systems are interconnected in such a manner that one provides a means of heat rejection for the other. Cascade systems are desirable for applications requiring temperatures between -50 and -150°F and allow the use of different refrigerants in each cycle (Green, 2008). Theoretically, any number of cascaded stages are possible, each stage requiring an additional condenser and an additional stage of compression.

In refrigerated condenser systems, two kinds of refrigerants are used, primary and secondary. Primary refrigerants are those that undergo a phase change from liquid to gas after absorbing heat. Examples are ammonia (R-717), liquid nitrogen (R-728), chlorofluorocarbons, hydrochlorofluorocarbons such as chlorodifluoromethane (R-22), or hydrofluorocarbons such as

R-410A or 1,1,1,2-Tetrafluoroethane (R-134a). Concerns about CFCs and HCFCs causing depletion of the ozone layer prompted development of HFCs, and recent concerns about HFCs being potent greenhouse gases are further prompting development of systems using alternate refrigerants (EPA, 2001, and EPA, 2015).

Secondary refrigerants such as brine solutions act only as heat carriers and remain in liquid phase. Conventional systems use a closed primary refrigerant loop that cools the secondary loop through the heat transfer medium in the evaporator. The secondary heat transfer fluid is then pumped to a VOC vapor condenser where it is used to cool the air/VOC vapor stream. In some applications, however, the primary refrigeration fluid is directly used to cool the vapor stream.

2.2.3 Auxiliary Equipment

As shown in Figure 2.2, some applications may require auxiliary equipment such as precoolers, recovery/storage tanks, pumps/blowers, and piping.

If water vapor is present in the treated gas stream or if the VOC has a high freezing point (e.g., benzene), ice or frozen hydrocarbons may form on the condenser tubes or plates. This will reduce the heat transfer efficiency of the condenser and thereby reduce the removal efficiency. Formation of ice will also increase the pressure drop across the condenser. In such cases, a precooler may be needed to condense the moisture prior to the VOC condenser. This precooler would bring the temperature of the stream down to approximately 35 to 40°F, effectively removing the moisture from the gas. Alternatively, an intermittent heating cycle can be used to melt away ice build-up. This may be accomplished by circulating ambient temperature brine through the condenser or by the use of radiant heating coils. If a system is not operated continuously, the ice can also be removed by circulating ambient air.

A VOC recovery tank for temporary storage of condensed VOC prior to reuse, reprocessing, or transfer to a larger storage tank may be necessary in some cases. Pumps and blowers are typically used to transfer liquid (e.g., coolant or recovered VOC) and gas streams, respectively, within the system.

2.3   Design Procedures

In this section, two procedures are presented for designing (sizing) refrigerated surface condenser systems to remove VOC from air/VOC mixtures. With the first procedure presented, one calculates the condenser exit temperature needed to obtain a given VOC recovery efficiency. In the second procedure, which is the inverse of the first, the exit temperature is given and the recovery efficiency corresponding to it is calculated.

The first procedure depends on knowledge of the following parameters:

1. Volumetric flow rate of the VOC-containing gas stream;

2. Inlet temperature of the gas stream;

3. Concentration and composition of the VOC in the gas stream;

4. Required removal efficiency of the VOC;

5. Moisture content of the emission stream; and

6. Properties of the VOC (assuming the VOC is a pure compound):

   - Heat of condensation,
   - Heat capacity, and
   - Vapor pressure.

The design of a refrigerated condenser system requires determination of the VOC condenser size and the capacity of the refrigeration unit. For a given VOC removal efficiency, the condensation temperature and the heat load need to be calculated to determine these parameters. The data necessary to perform the sizing procedures as well as the variable names and their respective units are listed in Table 2.1.

Table 2.1: Required Input Data

Data	Variable Name	Units
Inlet Stream Flow Rate	Q_{in}	scfm (at 77 °F; 1 atm)
Inlet Stream Temperature	T_{in}	°F
VOC Inlet Volume Fraction	$y_{VOC,in}$	Volume fraction
Required VOC Removal Efficiency	η	Fractional (volume)
Antoine Equation Constants[a]	A, B, C	—
Heat of Condensation of the VOC[a]	ΔH	Btu/lb-mole
Heat Capacity of the VOC[a]	$C_{p,VOC}$	Btu/lb-mole-°F
Specific Heat of the Coolant	$C_{p,cool}$	Btu/lb-°F
Heat Capacity of Air	$C_{p,air}$	Btu/lb-mole-°F

[a]See Appendix A for these properties of selected organic compounds.

The steps outlined below for estimating condensation temperature and the heat load apply to a two-component mixture (VOC/air) in which one of the two components is considered to be noncondensible (air). The VOC component is assumed to consist of a single compound. Also, the emission stream is assumed to be free of moisture. The calculations are based on the assumptions of ideal gas and ideal solution to simplify the sizing procedures. For a more rigorous analysis, see Reference Kern, 1950.

2.3.1   Estimating Condensation Temperature

The temperature necessary to condense the required amount of VOC must be estimated to determine the heat load. The first step is to determine the VOC concentration at the outlet of the condenser for a given removal efficiency. This is calculated by first determining the partial pressure of the VOC at the outlet, P_{VOC}. Assuming that the ideal gas law applies, P_{VOC} is given by:

$$P_{VOC} = 760 \times \frac{M_{VOC,out}}{M_{in} - M_{VOC,recovered}} \tag{2.1}$$

where:

P_{VOC} = partial pressure of the VOC in the exit stream (mm Hg)
M_{in} = moles in the inlet stream (moles per hour, moles/hr)
$M_{VOC,out}$ = moles of VOC in the outlet stream (moles/hr)
$M_{VOC,recovered}$ = moles of VOC condensed, or recovered (moles/hr)

The condenser is assumed to operate at a constant pressure of one atmosphere (760 mm Hg).

However:

$$M_{VOC,out} = M_{VOC,in} \times (1 - \eta) \tag{2.2}$$

where:

$M_{VOC,in}$ = moles of VOC in the inlet stream (moles/hr)
η = removal efficiency of the condenser system (fractional)

and

$$M_{VOC,in} = M_{in} \times y_{VOC,in} \tag{2.3}$$

where:

$y_{VOC,in}$ = volume fraction of VOC in inlet stream.

The removal efficiency, η, can also be defined as "moles VOC recovered / moles VOC in inlet:"

$$\eta = \frac{moles\ VOC\ recovered}{moles\ VOC\ in\ inlet}$$

or:

$$\eta = \frac{M_{VOC,recovered}}{M_{VOC,in}} \tag{2.4}$$

Rearranging Equation 2.4, we obtain:

$$M_{VOC,recovered} = M_{VOC,in} \times \eta \tag{2.5}$$

After substituting these variables in Equation 2.1, we obtain:

$$P_{VOC} = 760 \times \left[\frac{y_{VOC,in} \times (1 - \eta)}{[1 - (\eta \times y_{VOC,in})]} \right] \qquad (2.6)$$

At the condenser outlet, the VOC in the gas stream is assumed to be at equilibrium with the VOC condensate. At equilibrium, the partial pressure of the VOC in the gas stream is equal to its vapor pressure at that temperature assuming the condensate is pure VOC (i.e., vapor pressure P_{VOC}). Therefore, by determining the temperature at which this condition occurs, the condensation temperature can be specified. This calculation is based on the Antoine equation that defines the relationship between vapor pressure and temperature for a particular compound:

$$\log_{10}(P_{VOC}) = A - \frac{B}{T_{con} + C} \qquad (2.7)$$

where:

T_{con} = condensation temperature (degrees Celsius, or °C)
A, B, C = VOC-specific constants pertaining to temperature expressed in °C and pressure in millimeters of mercury (mm Hg) (see Appendix A, Table A-2)

Rearranging and solving for T_{con} in degrees Fahrenheit (ºF):

$$T_{con} = \left(\frac{B}{A - \log_{10}(P_{VOC})} - C \right) \times 1.8 + 32 \qquad (2.8)$$

The calculation methods for a gas stream containing multiple VOCs are complex, particularly when there are significant departures from the ideal behavior of gases and liquids. However, the temperature necessary for condensation of a mixture of VOCs can be estimated by the weighted average of the temperatures necessary to condense each VOC in the gas stream at a concentration equal to the total VOC concentration (Erikson, 1980).

2.3.2   VOC Condenser Heat Load

Condenser heat load is the amount of heat that must be removed from the inlet stream to attain the specified removal efficiency. It is determined from an energy balance, taking into account the enthalpy change due to the temperature change of the VOC, the enthalpy change due to the condensation of the VOC, and the enthalpy change due to the temperature change of the air. Enthalpy change due to the presence of moisture in the inlet gas stream is neglected in the following analysis.

For the purpose of this estimation, it is assumed that the total heat load on the system is equal to the VOC condenser heat load. Realistically, when calculating refrigeration capacity requirements for low temperature cooling units, careful consideration should be given to the process line losses and heat input of the process pumps. Refrigeration unit capacities are typically rated in terms of net output and do not reflect any losses through process pumps or process lines.

First, the number of lb-moles of VOC per hour in the inlet stream must be calculated by the following expression:

$$M_{VOC,in} = \frac{Q_{in}}{392} \times \left(y_{VOC,in}\right) \times 60 \tag{2.9}$$

where:

$M_{VOC,in}$ = moles of VOC in the inlet stream, lb-moles/hr
Q_{in} = volume flow rate in standard ft³/min (scfm)
392 = volume (ft³) occupied by one lb-mole of gas at standard conditions (77°F and 1 atm)
60 = conversion from minutes to hours (min/hr)

The number of lb-moles of VOC per hour in the outlet gas stream is calculated using Equation 2.2. Finally, the number of lb-moles of VOC per hour that are condensed (or recovered) is calculated as follows:

$$M_{VOC,recovered} = M_{VOC,in} - M_{VOC,out} \tag{2.10}$$

The condenser heat load is calculated by the following equation:

$$H_{load} = \Delta H_{con} + \Delta H_{uncon} + \Delta H_{noncon} \tag{2.11}$$

where:

H_{load} = condenser heat load (Btu/hr)
ΔH_{con} = enthalpy change associated with the condensed VOC (Btu/hr)
ΔH_{uncon} = enthalpy change associated with the uncondensed VOC (Btu/hr)
ΔH_{noncon} = enthalpy change associated with the noncondensible air (Btu/hr)

The change in enthalpy of the condensed VOC is calculated as follows:

$$\Delta H_{con} = M_{VOC,recovered} \left[\Delta H_{VOC} + C_{p,VOC}\left(T_{in} - T_{con}\right)\right] \tag{2.12}$$

where:

ΔH_{VOC} = molar heat of condensation of the VOC (Btu/lb-mole)
$C_{p,VOC}$ = molar heat capacity of the VOC (Btu/lb-mole-°F)
T_{in} = inlet stream temperature (°F)
T_{con} = condensation temperature (°F)

Each parameter varies as a function of temperature. In Equation 2.12, ΔH_{VOC} and $C_{p,VOC}$ are evaluated at the mean temperature:

$$T_{mean} = \frac{T_{in} + T_{con}}{2} \tag{2.13}$$

In general, the heat of condensation at a specific temperature, T_2 (degrees Rankine, or °R), can be calculated from the heat of condensation at a reference temperature, T_1 (°R), using the Watson Equation (Poling, Prausnitz, and O'Connell, 2001):

$$\left(\Delta H_{VOC} \ at \ T_2\right) = \left(\Delta H_{VOC} \ at \ T_1\right)\left(\frac{1 - \dfrac{T_2}{T_c}}{1 - \dfrac{T_1}{T_c}}\right)^{0.38} \tag{2.14}$$

where T_c (°R) is the VOC critical temperature.

The heat capacity can also be calculated for a specific temperature, T_2, if heat capacity constants are known for the particular compound. The heat capacity equation is:

$$C_{p,VOC} = a + bT_2 + cT_2^{\,2} + dT_2^{\,3} \tag{2.15}$$

where:

   a, b, c, d = compound-specific heat capacity constants

However, to simplify the heat load analysis, $C_{p,VOC}$ can be assumed to remain constant over the temperature range of operation (i.e., $T_{in} - T_{con}$) without much loss of accuracy in the heat load calculations, as the sensible heat change in Equation 2.12 is relatively small compared to the enthalpy change due to condensation.

Heat of condensation and heat capacity data are provided in Appendix A, Table A-1, along with critical temperatures. The heat of condensation for each compound is reported at its boiling point, while its heat capacity is given at 77 °F. To estimate the heat of condensation at another temperature, use Equation 2.14. However, the Appendix A heat capacity data may be used to approximate $C_{p,VOC}$ at other temperatures, since sensible heat changes are usually small, compared to condensation enthalpy changes.

The enthalpy change associated with the uncondensed VOC is calculated by the following expression:

$$\Delta H_{uncon} = M_{VOC,out} \, C_{p,VOC} \, (T_{in} - T_{con}) \tag{2.16}$$

Finally, the enthalpy change of the noncondensible air is calculated as follows:

$$\Delta H_{noncon} = \left[\left(\frac{Q_{in}}{392} \times 60 \right) - M_{VOC,in} \right] C_{p,air} (T_{in} - T_{con}) \qquad (2.17)$$

where $C_{p,air}$ is the specific heat of air. In both Equations 2.16 and 2.17, the C_p's are evaluated at the mean temperature, as given by Equation 2.13.

2.3.3 Condenser Size

Condensers are sized based on the heat load, the logarithmic mean temperature difference between the emission and coolant streams, and the overall heat transfer coefficient. The overall heat transfer coefficient, U, can be estimated from individual heat transfer coefficients of the gas stream and the coolant. The overall heat transfer coefficients for tubular heat exchangers where organic solvent vapors in noncondensible gas are condensed on the shell side and water/brine is circulated on the tube side typically range from 20 to 60 Btu/hr-ft²-°F according to *Perry's Chemical Engineers' Handbook* (Green, 2008). To simplify the calculations, a single U value may be used to size these condensers. This approximation is acceptable for purposes of making study-level (±30% accuracy) cost estimates.

Accordingly, an estimate of 20 Btu/hr-ft²-°F can be used to obtain a conservative estimate of condenser size (i.e., more likely to overstate the required condenser surface area). The following equation is used to determine the required heat transfer area:

$$A_{con} = \frac{H_{load}}{U \, \Delta T_{lm}} \qquad (2.18)$$

where:

A_{con} = condenser surface area (ft²)
U = overall heat transfer coefficient (Btu/hr-ft²-°F)
ΔT_{lm} = logarithmic mean temperature difference (°F).

While not all condensing units will experience fouling, for some applications of condensers, future, continued operation may result in fouling of the condenser surface areas, and facilities often account for this in the original condenser design by including a fouling factor in the overall heat transfer coefficient.[2, 3] Because fouling reduces heat transfer, use of a fouling factor reduces the overall heat transfer coefficient and thereby increases the surface area estimated for the condensing unit. This overall heat transfer coefficient that accounts for fouling is sometimes referred to as "U_{dirty}" or "Dirty U." The fouling factor depends on the materials of construction, the VOC condensed and other pollutants present, and the type of coolant used. The

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[2] Ulhrich provides the overall heat transfer coefficient equation, including use of fouling coefficients. Uhlrich, G., 1984. *A Guide to Chemical Engineering Process Design and Economics*. Wiley & Sons, Inc., New York.
[3] Perry's provides typical overall heat transfer coefficients, see Table 10-10. Perry's, 1984. *Perry's Chemical Engineers' Handbook, 6th Edition*. McGraw-Hill Book Company, New York.

use of relatively lower values of U, such as the estimate for Eqn. 2.18 above, could provide some allowance for fouling in the design and cost estimation for condensers.

The logarithmic mean temperature difference is calculated by the following equation, which is based on the use of a countercurrent flow condenser:

$$\Delta T_{lm} = \frac{(T_{in} - T_{cool,out}) - (T_{con} - T_{cool,in})}{\ln\left(\dfrac{T_{in} - T_{cool,out}}{T_{con} - T_{cool,in}}\right)} \qquad (2.19)$$

where:

$T_{cool,in}$ = coolant inlet temperature (°F)
$T_{cool,out}$ = coolant outlet temperature (°F).

The temperature difference ("approach") at the condenser exit can be assumed to be 15°F. In other words, the coolant inlet temperature, $T_{cool,in}$, will be 15°F less than the calculated condensation temperature, T_{con}. Also, the temperature rise of the coolant is specified as 25°F. (These two temperatures—the condenser approach and the coolant temperature rise—reflect good design practice that, if used, will result in an acceptable condenser size.) Therefore, the following equations can be applied to determine the coolant inlet and outlet temperature:

$$T_{cool,in} = T_{con} - 15\,°F \qquad (2.20)$$

$$T_{cool,out} = T_{cool,in} + 25\,°F \qquad (2.21)$$

2.3.4 Coolant Flow Rate

The heat removed from the emission stream is transferred to the coolant. By a simple energy balance, the flow rate of the coolant can be calculated as follows:

$$W_{cool} = \frac{H_{load}}{C_{p,cool} \times (T_{cool,out} - T_{cool,in})} \qquad (2.22)$$

where:

W_{cool} = coolant flow rate (lb/hr)
$C_{p,cool}$ = coolant specific heat (Btu/lb-°F)

$C_{p,cool}$ will vary according to the type of coolant used. For a 50:50 (volume %) mixture of ethylene glycol and water, $C_{p,cool}$ is approximately 0.65 Btu/lb-°F. The specific heat of brine (salt water), another commonly used coolant, is approximately 1.0 Btu/lb-°F.

2.3.5 Refrigeration Capacity

The refrigeration unit is assumed to supply the coolant at the required temperature to the condenser. The required refrigeration capacity is calculated by converting the condenser heat load from Btu/hr to tons of refrigeration as follows:

$$R = \frac{H_{load}}{12,000} \qquad (2.23)$$

where:

    12,000 = conversion to refrigeration tons (Btu/ton)

Again, as explained in section 2.3.2, H_{load} does not include any heat losses.

2.3.6 Recovered VOC

The mass of VOC recovered in the condenser can be calculated using the following expression:

$$W_{VOC,recovered} = M_{VOC,recovered} \times MW_{VOC} \qquad (2.24)$$

where:

$W_{VOC,recovered} =$ mass of VOC recovered (or condensed) (lb/hr)
    $MW_{VOC} =$ molecular weight of the VOC (lb/lb-mole)

2.3.7 Auxiliary Equipment

As noted previously, the auxiliary equipment for a refrigerated surface condenser system may include:

- precooler,
- recovered VOC storage tank,
- pumps/blowers, and
- piping/ductwork.

If water vapor is present in the treated gas stream, a precooler may be needed to remove moisture to prevent ice from forming in the VOC condenser. Sizing of a precooler is influenced by the moisture concentration and the temperature of the emission stream. As discussed in Section 2.2.3, a precooler may not be necessary for intermittently operated refrigerated surface condenser systems where the ice will have time to melt between successive operating periods.

If a precooler is required, a typical operating temperature is 35 to 40°F. At this temperature, almost all of the water vapor present will be condensed without danger of freezing. These condensation temperatures roughly correspond to a removal efficiency range of 70 to 80

percent if the inlet stream is saturated with water vapor at 77°F. The design procedure outlined in the previous sections for a VOC condenser can be used to size a precooler, based on the psychrometric chart for the air-water vapor system (Green, 2008).

Storage/recovery tanks are used to store the condensed VOC when direct recycling is not a suitable option. The size of these tanks is determined from the amount of VOC condensate to be collected and the amount of time necessary before unloading. Sizing of pumps and blowers is based on the liquid and gas flow rates, respectively, as well as the system pressure drop between the inlet and outlet. Sizing of the piping and ductwork (length and diameter) primarily depends upon the stream flow rate, duct/pipe velocity, available space, and system layout.

2.3.8   Alternate Design Procedure

In some applications, it may be desirable to size and cost a refrigerated condenser system that will use a specific coolant and provide a particular condensation temperature. The design procedure to be implemented in such a case would essentially be the same as the one presented in this section except that instead of calculating the condenser exit temperature needed to obtain a specified VOC recovery efficiency, the exit temperature is given and the corresponding recovery efficiency is calculated.

The initial calculation would be to estimate the partial (i.e., vapor) pressure of the VOC at the given condenser exit temperature, T_{con}, using Equation 2.7. Next, calculate η using Equation 2.25, by rearranging Equation 2.6:

$$\eta = \frac{\left(760\ y_{VOC,in}\right) - P_{VOC}}{y_{VOC,in} \times \left(760 - P_{VOC}\right)} \tag{2.25}$$

Finally, substitute the calculated P_{VOC} into this equation to obtain η. In the remainder of the calculations to estimate condenser heat load, refrigeration capacity, coolant flow rate, etc., follow the procedure presented in Sections 2.3.2 through 2.3.7.

In future application and operation of the condenser, if process changes result in raw material, product, or solvent VOC changes; VOC concentration changes; volumetric flowrate changes; or process temperature changes, the removal or recovery efficiency achieved may differ from the original VOC design efficiency.

2.4   Estimating Total Capital Investment

This section presents the procedures and data necessary for estimating capital costs for refrigerated surface condenser systems in solvent vapor recovery and gasoline vapor recovery applications. Costs for packaged and nonpackaged solvent vapor recovery systems are presented in Sections 2.4.1 and 2.4.2, respectively. Costs for packaged gasoline vapor recovery systems are described in Section 2.4.3. Costs are calculated based on the design/sizing procedures discussed in Section 2.3.

Total capital investment, *TCI*, includes equipment cost, *EC*, for the entire refrigerated condenser unit, auxiliary equipment costs, taxes, freight charges, instrumentation, and direct and

indirect installation costs. All costs in this chapter are presented in 2014 dollars (2014$). The costs were escalated from 1990 dollars using a ratio of the years' annual composite Chemical Engineering Plant Cost Indices (CEPCI). A ratio of annual CEPCI for 2014 and 1990 results in an escalation factor of 1.611 (i.e., 576.1/357.6) (Lozowski, 2015, and Vatavuk, 2002). Typically, it is more common to perform this kind of escalation when the difference between the years is small (i.e., 5 or less) because emerging technologies and practices can increase the inaccuracy of a simple ratio.

For these control systems, the total capital investment is a battery limit cost estimate and does not include the provisions for bringing utilities, services, or roads to the site; the backup facilities; the land; the working capital; the research and development required; or the process piping and instrumentation interconnections that may be required in the process generating the waste gas. These costs are based on new plant installations; no retrofit cost considerations are included. The retrofit cost factors are so site specific that no attempt has been made to provide them.

The expected accuracy of the cost estimates presented in this chapter is ±30 percent (i.e., "study" estimates). It must be kept in mind that even for a given application, design and manufacturing procedures vary from vendor to vendor, so costs may vary.

In the next two sections, equipment costs are presented for packaged and nonpackaged (custom) solvent vapor recovery systems, respectively. With the packaged systems, the equipment cost is factored from the refrigeration unit cost; with the custom systems, the equipment cost is determined as the sum of the costs of the *individual* system components. It should be noted that any adjustments for fouling, if made, will not affect the cost estimates for the packaged recovery systems. The cost equations for these systems are based on refrigeration and temperature as variables, and surface area is not a variable. Surface area is a variable in the custom systems equations, and effects from fouling are explicitly accounted for. Finally, equipment costs for packaged gasoline vapor recovery systems are given in Section 2.4.3.

2.4.1   Equipment Costs for Packaged Solvent Vapor Recovery Systems

Vendors were asked to provide refrigerated unit cost estimates for a wide range of applications. The equations shown below for refrigeration unit equipment costs, EC_r, are multivariable regressions of data provided by two vendors and are only valid for the ranges listed in Table 2.2 (Sisk, 1991, and Waldrop and Sardo, 1990). In this table, the capacity range of refrigeration units for which cost data were available are shown as a function of temperature.

Table 2.2: Applicability Ranges for the Refrigeration Unit Cost Equations (Equations 2.26 to 2.28)

Temperature T_{con} (°F)[a]	Minimum Size Available R (tons)		Maximum Size Available R (tons)	R (tons)
	Single Stage	Multistage	Single Stage	Multistage
40	0.85	NA[b]	174	NA
30	0.63	NA	170	NA
20	0.71	NA	880	NA
10	0.44	NA	200	NA
0 to -5	0.32	NA	133	NA
-10	0.21	3.50	6.6	81
-20 to -25	0.13	2.92	200	68
-30	NA	2.42	NA	85
-40	NA	1.92	NA	68
-45 to -50	NA	1.58	100[c]	55
-55 to -60	NA	1.25	100[c]	100
-70	NA	1.33	NA	42
-75 to -80	NA	1.08	NA	150
-90	NA	0.83	NA	28
-100	NA	0.67	NA	22

[a] For condensation temperatures that lie between the levels shown, round off to the nearest level (e.g., if T_{con} = 16 °F, use 20 °F) to determine minimum and maximum available size.
[b] NA = System not available based on vendor data collected in this study.
[c] Only one data point available.

Single Stage Refrigeration Units (less than 10 tons)

$$EC_r = 1.611 \times \exp\left(9.83 - 0.014\, T_{con} + 0.340\, \ln R\right) \tag{2.26}$$

Single Stage Refrigeration Units (greater than or equal to 10 tons)

$$EC_r = 1.611 \times \exp\left(9.26 - 0.007\, T_{con} + 0.627\, \ln R\right) \tag{2.27}$$

Multistage Refrigeration Units

$$EC_r = 1.611 \times \exp\left(9.73 - 0.012\, T_{con} + 0.584\, \ln R\right) \tag{2.28}$$

(NOTE: $\exp(a) = e^a \approx 2.718^a$)

Equations 2.26 and 2.27 provide costs for refrigeration units based on single stage designs, while Equation 2.28 gives costs for multistage units. Equation 2.28 covers both types of multistage units, "cascade" and "compound." Data provided by a vendor show that the costs of cascade and compound units compare well, generally differing by less than 30% (Sisk, 1991). Thus, only one cost equation is provided. Equation 2.26 applies to single stage refrigeration units smaller than

10 tons and Equation 2.27 applies to single stage refrigeration units as large as or larger than 10 tons. Single stage units typically achieve temperatures between 40 and -20°F, although there are units that are capable of achieving -60°F in a single stage (Sisk, 1991, and Price, 1991). Multistage units are capable of lower temperature operation between -10 and -100°F.

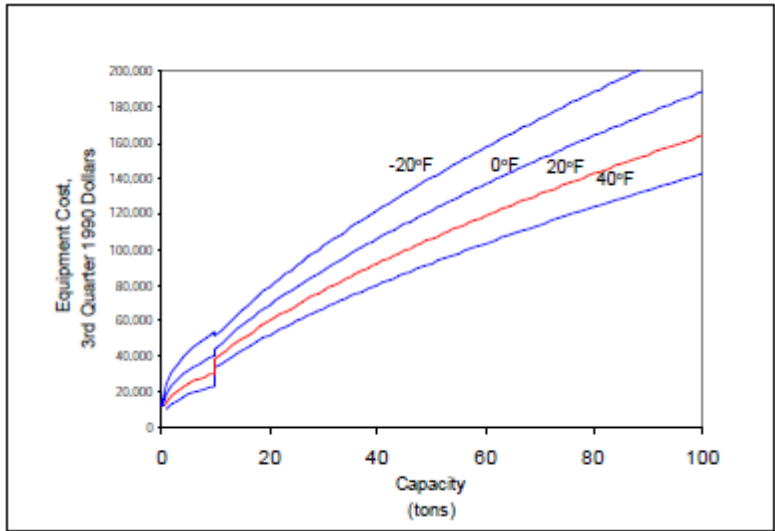**Figure 2.5:** Refrigeration Unit Equipment Cost (Single Stage) (Sisk, 1991, and Waldrop and Sardo, 1990)

Single stage refrigeration unit costs are depicted graphically for selected temperatures in Figure 2.5. Figure 2.6 shows the equipment cost curves for multistage refrigeration units.
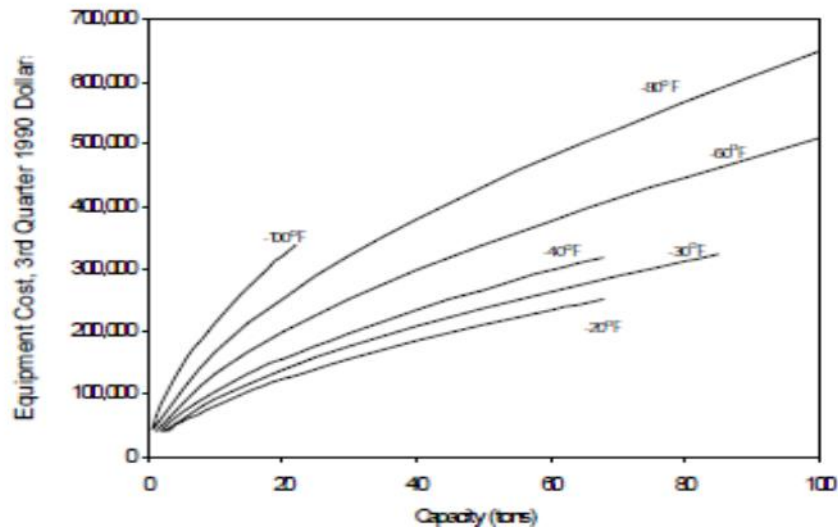

Figure 2.6: Refrigeration Unit Equipment Cost (Multistage) (Sisk, 1991, and Waldrop and Sardo, 1990)

(NOTE: In Figure 2.5, the discontinuities in the curves at the 10 ton capacity are a result of the two regression equations used. Equation 2.26 is used for refrigeration capacities of less than 10 tons; Equation 2.27 is used for capacities greater than or equal to 10 tons.)

These costs are for outdoor models that are skid-mounted on steel beams and consist of the following components: walk-in weatherproof enclosure, air-cooled low temperature refrigeration machinery with dual pump design, storage reservoir, control panel and instrumentation, vapor condenser, and necessary piping. All refrigeration units have two pumps: a system pump and a bypass pump to short-circuit the vapor condenser during no-load conditions. Costs for heat transfer fluids (brine) are not included.

The equipment cost of packaged solvent vapor recovery systems (EC_p) is estimated to be 25 percent greater than the cost of the refrigeration unit alone (Waldrop and Sardo, 1990). The additional cost includes VOC condenser, recovery tank, the necessary connections, piping, and additional instrumentation. Thus,

$$EC_p = 1.25 \, EC_r \qquad (2.29)$$

Purchased equipment cost, PEC_p, includes the packaged equipment cost, EC_p, and factors for sales taxes (0.03) and freight (0.05). Instrumentation and controls are included with the packaged units. Thus,

$$PEC_p = EC_p \left(1 + 0.03 + 0.05\right) = 1.08 \, EC_p \qquad (2.30)$$

2.4.2 Equipment Costs for Nonpackaged (Custom) Solvent Vapor Recovery Systems

To develop cost estimates for nonpackaged or custom refrigerated systems for the previous Manual chapter (Sixth Edition) on condensers, information was solicited from vendors on costs of refrigeration units, VOC condensers, and VOC storage/recovery tanks (Waldrop and Sardo, 1990, Hansek, 1990, and Cooke, 1990). Quotes from the vendors were used to develop the estimated costs for each type of equipment. Only one set of vendor data was available for each type of equipment.

Equations 2.26, 2.27, and 2.28 shown above are applicable for estimating the costs for the refrigeration units, EC_r.

Equation 2.31 shows the equation developed for the VOC condenser cost estimates (Hansek, 1990):

$$EC_{con} = 1.611 \times \left(34 \, A_{con} + 3,775\right) \qquad (2.31)$$

This equation is valid for the surface area range of 38 to 800 ft^2 and represents costs for shell and tube type heat exchangers with 304 stainless steel tubes.

The following equation represents the storage/recovery tank cost data obtained from one vendor (Cooke, 1990):

$$EC_{tank} = 1.611 \times (2.72\ V_{tank} + 1{,}960)$$ (2.32)

These costs are applicable for the storage tank volume range of 50 to 5,000 gallons and pertain to 316 stainless steel vertical tanks.

Costing procedures for a precooler (EC_{pre}) that includes a separate condenser/refrigeration unit and a recovery tank are similar to that for a custom refrigerated condenser system. Hence, Equations 2.26 through 2.32 would be applicable, with the exception of Equation 2.28, which represents multistage systems. Multistage systems operate at much lower temperatures than that required by a precooler.

Costs for auxiliary equipment such as ductwork, piping, fans, or pumps are designated as EC_{aux}, These items should be costed separately using methods described elsewhere in this Manual.

The total equipment cost for custom systems, EC_c is then expressed as:

$$EC_c = EC_r + EC_{con} + EC_{tank} + EC_{pre} + EC_{aux}$$ (2.33)

The purchased equipment cost including EC_c and factors for sales taxes (0.03), freight (0.05), and instrumentation and controls (0.10) is given below:

$$PEC_c = EC_c(1 + 0.03 + 0.05 + 0.10) = 1.18\ EC_c$$ (2.34)

2.4.3   Equipment Costs for Gasoline Vapor Recovery Systems

Separate quotes were obtained for packaged gasoline vapor recovery systems because these systems are specially designed for controlling gasoline vapor emissions from such sources as storage tanks, gasoline bulk terminals, and marine vessel loading and unloading operations. Systems that control marine vessel gasoline loading and unloading operations also must meet U.S. Coast Guard safety requirements.

Quotes obtained from one vendor were used to develop equipment cost estimates for these packaged systems (see Figure 2.7). The cost equation shown below is a least squares regression of these cost data and is valid for the refrigeration capacity range of 20 to 140 tons (Waldrop and Sardo, 1990).

$$EC_p = 1.611 \times (4{,}910\ R + 341{,}000)$$ (2.35)

The vendor data in process flow capacity (gal/min) versus cost ($) were transformed into Equation 2.35 after applying the design procedures in Section 2.3. Details of the data transformation are given in Appendix B.

The cost estimates apply to skid-mounted refrigerated VOC condenser systems for hydrocarbon vapor recovery primarily at gasoline loading/storage facilities. The systems are intermittently operated at -80 to -120°F allowing 30 to 60 minutes per day for defrosting by

circulation of warm brine. Multistage systems are employed to achieve these lower temperatures. The achievable VOC removal efficiencies for these systems are in the range of 90 to 95 percent.

The packaged systems include the refrigeration unit with the necessary pumps, compressors, condensers/evaporators, coolant reservoirs, the VOC condenser unit and VOC recovery tank, precooler, instrumentation and controls, and piping. Costs for heat transfer fluids (brines) are not included. The purchased equipment cost for these systems includes sales tax and freight and is calculated using Equation 2.30.
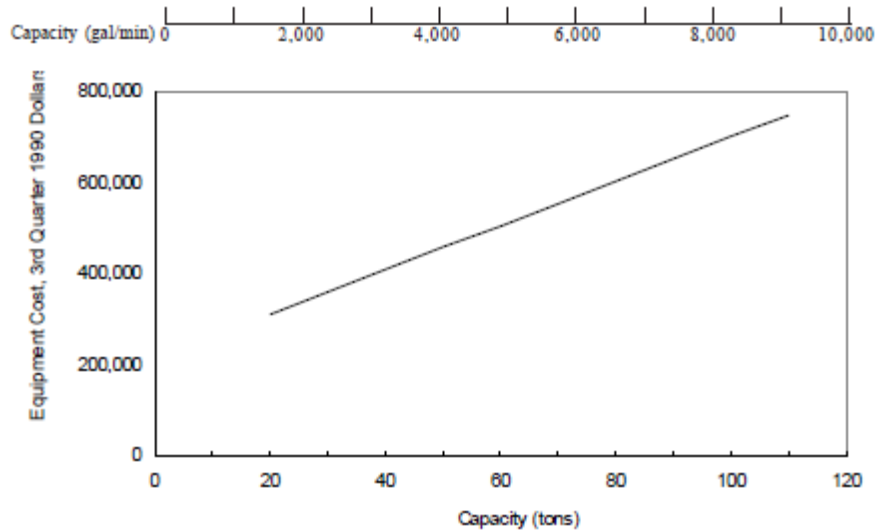
Figure 2.7: Gasoline Vapor Recovery System Equipment Cost (Waldrop and Sardo, 1990)

2.4.4   Installation Costs

The total capital investment, *TCI*, for packaged systems is obtained by multiplying the purchased equipment cost, PEC_p by the total installation factor (Waldrop, 1988):

$$TCI = 1.15\,PEC_p \qquad (2.36)$$

For nonpackaged (custom) systems, the total installation factor is 1.71:

$$TCI = 1.71\,PEC_c \quad (2.37)$$

An itemization of the total installation factor for nonpackaged systems is shown in Table 2.3. Depending on the site conditions, the installation costs for a given system could deviate significantly from costs generated by these average factors, as well as from additional costs incurred from site preparation, additional buildings and other unexpected costs. Costs for site preparation (SP), buildings (Bldg.), and contingencies (C) must be included in the TCI for

custom systems, as shown in Table 2.3. Guidelines are available for adjusting these average installation factors (Vatavuk and Neveril, 1980).

Table 2.3: Capital Cost Factors for Nonpackaged (Custom) Refrigerated Condenser Systems

Cost Item	Factor
Direct Capital Costs	
Purchased Equipment Costs	
Refrigerated condenser system, EC	As estimated, A
Instrumentation	0.10 A
Sales Taxes	0.03 A
Freight	0.05 A
Purchased equipment costs, PEC	$B = 1.18\ A^a$
Direct Capital Costs (Installation)	0.08 B
Foundations & Support	0.14 B
Handling & Erections	0.08 B
Electrical	0.08 B
Piping	0.02 B
Insulation	0.10 B
Painting	0.01 B
Direct Costs (Installation)	0.43 B
Site Preparation	As required, SP
Buildings	As required, Bldg.
Total Direct Costs, DC	1.43 B + SP + Bldg.
Indirect Capital Costs (Installation)	
Engineering	0.10 B
Construction and Field Expenses	0.05 B
Contractor Fees	0.10 B
Start-Up	0.02 B
Performance Test	0.01 B
Total Indirect Costs, IC	0.28 B
Contingency Costs (C)	CF(DC+IC)
Total Capital Investment (TCI) = DC + IC + C	$1.71\ B + SP + Bldg. + C^b$

[a] Purchased equipment cost factor for packaged systems is 1.18 with instrumentation included.

[b] For comparison, for packaged systems, total capital investment = $1.15 PEC_p$.

[d] The default value for the contingency factor, CF, is 0.10. However, values of between 0.05 and 0.15 may be included to account for unexpected costs associated with the fabrication and installation of the control system. More information on contingencies can be found in the cost estimation chapter of this Manual.

2.5   Estimating Total Annual Cost

The total annual cost, *TAC*, is the sum of the direct and indirect annual costs. The bases used in calculating annual cost factors are given in Table 2.4.

Table 2.4: Suggested Annual Cost Factors for Refrigerated Condenser Systems

Cost Item	Factor
Direct Annual Cost, DAC	
Operating Labor	
Operator	½ hour per shift
Supervisor	15% of operator
Maintenance	
Labor	½ hour per shift
Material	100% of maintenance labor
Electricity	See Table 2.5
Indirect Annual Costs, IAC	
Overhead	60% of total labor and maintenance material costs
Administrative Charges	2% of total Capital Investment
Property Tax	1% of Total Capital Investment
Insurance	1% of Total Capital Investment
Capital Recovery[a]	$0.0915 \times$ Total Capital Investment
Recovery Credits, RC	
Recovered VOC Credit	Quantity recovered $\times$ operating hours $\times$ Credit$_{VOC}$[b]
Total Annual Cost	DAC + IAC - RC

[a] See Section 1, Chapter 2 of the Control Cost Manual.. The capital recovery factor, CRF, is function of the refrigerated condenser equipment life and the opportunity cost of the capital (i.e., interest rate). For example, for a 15 year equipment life and a 4.25% interest rate, CRF = 0.0915.
[b] The value for recovered VOC, Credit$_{VOC}$, is in $/lb.

2.5.1   Direct Annual Costs

Direct annual costs, *DAC*, include labor (operating and supervisory), maintenance (labor and materials), and electricity. If fluorocarbon refrigerants are used in the condenser system, replacement costs of the refrigerant would also be DAC; costs for these should reflect the current costs for the calendar year of the analysis.[4] Operating labor is estimated at 1/2-hour per 8-hour shift. The supervisory labor cost is estimated at 15% of the operating labor cost. Maintenance

---

[4] Example costs for fluorocarbons in calendar year 2017: R22 (HCFC-22) are approximately $26/lb; R134a (HFC-134a) range from $3.70 to $6.60/lb; R410a (HFC-410a) are $9.20 to $10.40/lb; HFO-1234yf are $70/lb; R404a range from $3.50 to $10.40/lb; and R728 ($N_2$) are $0.17/lb.

labor is estimated at 1/2-hour per 8-hour shift. Maintenance materials costs are assumed to equal maintenance labor costs.

Utility costs for refrigerated condenser systems include electricity requirements for the refrigeration unit and any pumps/blowers. The power required by the pumps/blowers is negligible when compared with the refrigeration unit power requirements. Electricity requirements for refrigerated condenser systems are summarized in Table 2.5.

Table 2.5: Electricity Requirements

Temperature (°F)	Electricity (E, kW/ton)
40	1.3
20	2.2
-20	4.7
-50	5.0
-100	11.7

These estimates were developed from product literature obtained from one vendor (Waldrop and Sardo, 1990). The electricity cost, C_e, can then be calculated from the following expression:

$$C_e = \frac{R}{\eta_{compressor}} \times E \times \theta_s \times p_e \qquad (2.38)$$

where:

$$
\begin{aligned}
C_e &= \text{ electricity cost (\$/yr)} \\
\eta_{compressor} &= \text{ mechanical efficiency of the compressor} \\
E &= \text{ electricity requirement for the temperature achieved (kW/ton of refrigeration)} \\
\theta_s &= \text{ annual system operating hours (hr/yr)} \\
p_e &= \text{ electricity cost (\$/kW)}
\end{aligned}
$$

2.5.2 Indirect Annual Costs

Indirect annual costs, *IAC*, are calculated as the sum of capital recovery costs plus general and administrative (G&A), overhead, property tax, and insurance costs. Overhead is assumed to be equal to 60 percent of the sum of operating, supervisory, and maintenance labor, and maintenance materials. Overhead cost is discussed in Section 1 of this Manual.

The system capital recovery cost, *CRC*, is based on an estimated 15-year equipment life (Waldrop, 1988). (See Section 1 of the Manual for a discussion of the capital recovery cost.) The system capital recovery cost is then estimated by:

$$CRC = CRF \times TCI \qquad (2.39)$$

For a 15-year life and an interest rate of 4.25 percent, the capital recovery factor is 0.0915. G&A costs, property tax, and insurance are factored from total capital investment, typically at 2 percent, 1 percent, and 1 percent, respectively.

2.5.3 Recovery Credit

The recovered VOC may be disposed of or may be reused or sold. If the condensed VOC can be directly reused or sold without further treatment, then the credit from this operation is incorporated in the total annual cost estimates. The following equation can be used to estimate the VOC recovery credit, RC:

$$RC = W_{VOC,recovered} \times Credit_{VOC} \times \theta_s \qquad (2.40)$$

where:

$$
\begin{aligned}
RC &= \text{VOC recovery credit (\$/yr)} \\
Credit_{VOC} &= \text{resale value of recovered VOC (\$/lb)} \\
W_{VOC,recovered} &= \text{quantity of VOC recovered (lb/hr)} \\
\theta_s &= \text{annual hours of operation for the condenser (hr/yr)}
\end{aligned}
$$

2.5.4 Total Annual Cost

The total annual cost, TAC, is calculated as the sum of the direct and indirect annual costs, minus the recovery credit:

$$TAC = DAC + IAC - RC \qquad (2.41)$$

2.6 Example Problem 1

The example problem described in this section shows how to apply the refrigerated condenser system sizing and costing procedures to the control of a vent stream consisting of acetone, air, and a negligible amount of moisture. This example problem assumes a required removal efficiency and calculates the temperature needed to achieve this level of control.

2.6.1 Required Information for Design

The first step in the design procedure is to specify the gas stream to be processed. Gas stream parameters to be used in this example problem are listed in Table 2.6. The values for the Antoine equation constants, heat of condensation, and heat capacity of acetone are obtained from Appendix A. Specific heat of the coolant is obtained from *Perry's Chemical Engineers' Handbook* (Green, 2008).

Table 2.6: Example Problem Data

Vent Stream Parameters	Value
Inlet Stream Flow Rate	100 scfm[a]
VOC to be Condensed	Acetone
VOC Inlet Volume Fraction	0.375
Required VOC Removal Efficiency	0.90
Inlet Gas Stream Temperature	86 ºF
Antoine Equation Constants for Acetone:	
A	7.11714
B	1210.595
C	229.664
Heat of Condensation of Acetone[b]	12520 Btu/lb-mole
Heat Capacity of Acetone[c]	17.81 Btu/lb-mole-°F
Specific Heat of Coolant[c] (ethylene glycol)	0.65 Btu/lb-°F
Heat Capacity of Air[c]	6.95 Btu/lb-mole-°F
Annual Cost Parameters	**Value**
Operating Hours	2,080 hr/yr
Operating Labor[d]	$23.63/hr
Maintenance Labor[e]	$24.25/hr
Electricity[f]	$0.0701/kWh
Interest Rate	4.5%
Equipment Life	15 yrs.
Acetone Resale Value[g]	$0.26/lb

[a] Standard conditions: 77°F and 1 atmosphere.
[b] Evaluated at the acetone boiling point (134°F).
[c] These properties were evaluated at 77°F.
[d] Bureau of Labor Statistics, labor cost for "Chemical Operators" category, 2014
[e] Bureau of Labor Statistics, labor cost for "Industrial Mechanics" category, 2014
[f] U.S. EIA, "Electric Power Monthly", August 2015
[g] The Plaza Group, "Acetone Report", October 2015

2.6.2   Equipment Sizing

The first step in refrigerated condenser sizing is determining the partial pressure of the VOC at the outlet of the condenser for a given removal efficiency. Given the stream flow rate, inlet VOC concentration, and the required removal efficiency, the partial pressure of the VOC at the outlet can be calculated using Equation 2.6.

$$P_{VOC} = 760 \times \frac{0.375 \times (1 - 0.90)}{1 - 0.375 \times (0.90)} = 43.0 \ mm \ Hg$$

Next, the temperature necessary to condense the required amount of VOC must be determined using Equation 2.8:

$$T_{con} = \left(\frac{1210.595}{7.11714 - \log_{10}(43)} - 229.664 \right) 1.8 + 32 = 16°F$$

The next step is to estimate the VOC condenser heat load. Calculate: (1) the VOC flow rate for the inlet/outlet emission streams, (2) the flow rate of the condensed VOC, and (3) the condenser heat balance. The flow rate of VOC in the inlet stream is calculated from Equation 2.9.

$$M_{VOC,in} = \frac{100}{392} \times (0.375) \times 60 = 5.740 \frac{lb - moles}{hr}$$

The flow rate of VOC in the outlet stream is calculated using Equation 2.2 as follows:

$$M_{VOC,out} = 5.740 \times (1 - 0.90) = 0.5740 \frac{lb - moles}{hr}$$ Finally, the flow rate of condensed VOC is calculated with Equation 2.10:

$$M_{VOC,recovered} = 5.740 - 0.574 = 5.166 \frac{lb - moles}{hr}$$

Next, the condenser heat balance is conducted. As indicated in Table 2.6, the acetone heat of condensation is evaluated at its boiling point, 134°F. However, it is assumed (for simplicity) that all of the acetone condenses at the condensation temperature, $T_{con} = 16°F$. To estimate the heat of condensation at 16°F, use the Watson equation (Equation 2.14) with the following inputs:

T_c = 915 °R (from Appendix A)
T_1 = 134 + 460 = 594 °R
T_2 = 16 + 460 = 476 °R

Upon substitution, we obtain:

$$\left(\Delta H_{VOC} \ at \ 16 °F \right) = 12,520 \times \left(\frac{1 - \frac{476}{915}}{1 - \frac{594}{915}} \right)^{0.38}$$

$$= 14,090 \ \frac{Btu}{lb - mole}$$

As Table 2.6 shows, the heat capacities of acetone and air and the specific heat of the coolant were all evaluated at 77°F. This temperature is fairly close to the condenser mean operating temperature, i.e., (86 + 16)/2 = 51°F. Consequently, using the 77°F values would not add significant additional error to the heat load calculation.

The change in enthalpy of the condensed VOC is calculated using Equation 2.12:

$$\Delta H_{con} = 5.166 \times [14{,}090 + 17.81 \times (86 - 16)] = 79{,}230 \frac{Btu}{hr}$$

The enthalpy change associated with the uncondensed VOC is calculated from Equation 2.16:

$$\Delta H_{uncon} = (0.574) \times (17.81) \times (86 - 16) = 715.6 \frac{Btu}{hr}$$

Finally, the enthalpy change of the noncondensible air is estimated from Equation 2.17:

$$\Delta H_{noncon} = \left[\left(\frac{100}{392} \times 60 \right) - 5.740 \right] \times 6.95 \times (86 - 16) = 4{,}654 \frac{Btu}{hr}$$

The condenser heat load is then calculated by substituting ΔH_{con}, ΔH_{uncon}, and ΔH_{noncon} in Equation 2.11:

$$H_{load} = 79{,}230 + 715.6 + 4{,}654 = 84{,}600 \frac{Btu}{hr}$$

The next step is estimation of the VOC condenser size. The logarithmic mean temperature difference is calculated using Equation 2.19. In this calculation, from Equations 2.20 and 2.21, respectively:

$$T_{cool,in} \quad = \quad 16 - 15 = 1°F$$
$$T_{cool,out} \quad = \quad 1 + 25 = 26°F$$

and:

$$\Delta T_{lm} = \frac{(86 - 26) - (16 - 1)}{\left(\ln \dfrac{86 - 26}{16 - 1} \right)} = 32.5°F$$

The condenser surface area can then be calculated using Equation 2.18.

$$A_{con} = \frac{84{,}600}{20 \times (32.5)} = 130.2 \ ft^2$$

In this equation, a conservative value of 20 Btu/hr-ft^2-°F is used as the overall heat transfer coefficient.

The coolant flow rate can be calculated using Equation 2.22.

$$W_{cool} = \frac{84{,}600}{0.65 \times (26 - 1)} = 5{,}206 \ \frac{lb}{hr}$$

The refrigeration capacity can be estimated from Equation 2.23 as follows:

$$R = \frac{84{,}600}{12{,}000} = 7.05 \ tons$$

Finally, the quantity of recovered VOC can be estimated using Equation 2.24:

$$W_{VOC,re\,cov\,ered} = 5.166 \times 58.08 \ \frac{lb}{lb - mole} = 300.0 \ lb\,/\,hr$$

where the molecular weight of acetone is obtained from Appendix A.

Note that in this example case, the partial pressure of acetone at the condenser exit is relatively high (43 mm Hg). In applications where much lower outlet concentrations are desired, a second control device (e.g., incinerator, gas absorber, carbon adsorber) to operate in series with the condenser may need to be considered.

2.6.3 Equipment Costs

Once the system sizing parameters have been determined, the equipment costs can be calculated. For the purpose of this example, the costs of a custom refrigerated condenser system, including a refrigeration unit, a VOC condenser, and a recovery tank will be estimated.

From Table 2.2, a single stage refrigeration unit appears to be suitable for the example problem with an estimated condensation temperature of 16°F and capacity of 7.05 tons. Hence Equation 2.26, which is applicable to units less than 10 tons is selected for estimating costs. Application of this equation results in the following value for the refrigeration unit cost:

$$EC_r = 1.611 \times \exp\left[9.83 - 0.014\,(16) + 0.340\ \ln\,(7.05)\right] = \$46{,}480$$

VOC condenser cost is computed using Equation 2.31 as follows:

$$EC_{con} = 1.611 \times \left[34 \times (130\,.2) + 3{,}775\right] = \$13{,}210$$

Recovery tank cost can be calculated from Equation 2.32. For this case, $W_{VOC,recovered} =$ 300.0 lb/hr, which is equivalent to 45.45 gal/hr (density of acetone is about 6.6 lb/gal). Assuming an 8-hour daily operation, the interim storage capacity requirement would be 363.6 gallons. Application of Equation 2.32 leads to the following:

$$EC_{tank} = 1.611 \times \left[2.72 \times (363.64) + 1{,}960\right] = \$4{,}751$$

Assuming there are no additional costs due to precooler or other auxiliary equipment, the total equipment cost is calculated from Equation 2.33:

$$EC_c = 46,480 + 13,210 + 4,751 + 0 + 0 = \$64,440$$

The purchased equipment cost including instrumentation, controls, taxes, and freight is estimated using Equation 2.34:

$$PEC_c = 1.18 \times (64,440) = \$76,040$$

Using the default contingency factor of 0.10, the contingency cost (C) is estimated as follows:

$$C = 0.10 \times (1.71 \times 76,040) = \$13,000$$

The total capital investment is calculated using Equation 2.37:

$$TCI = 1.71 \times (76,040) + 13,000 = \$145,000$$

2.6.4   Total Annual Cost

Table 2.7 summarizes the estimated annual costs for the example problem. The cost calculations are shown in the table.

Direct annual costs for refrigerated systems include labor, materials, and utilities. Labor costs are based on 8-hr/day, 5-day/week operation. Supervisory labor is computed at 15 percent of operating labor, and operating and maintenance labor are each based on 1/2 hr per 8-hr shift. The electricity cost is based on a requirement of 2.2 kW/ton (see Table 2.5), because the condensation temperature (16°F) is close to the 20°F temperature given for this value.

Indirect annual costs include overhead, capital recovery, administrative charges, property tax, and insurance.

Total annual cost is estimated using Equation 2.41. For this example, application of refrigerated condensation as a control measure results in an annual savings of $122,000. As Table 2.7 shows, the acetone recovery credit is approximately four times the direct and indirect costs combined. Clearly, this credit has more influence on the total annual cost than any other component. Although the credit depends on three parameters—the acetone recovery rate, the annual operating hours, and the acetone salvage value ($0.26/lb) (Dunn, 2015) —the last parameter is often the most difficult to estimate. This is mainly because the salvage value varies according to the facility location and its local market as well as the current state of the broader chemical market.

Table 2.7: Annual Cost for Refrigerated Condenser System Example Problem

Cost Item	Calculations	Cost
Direct Annual Costs, DAC		
Operating Labor		
Operator	$\dfrac{0.5\ h}{shift} \times \dfrac{shift}{8\ h} \times \dfrac{2{,}080\ h}{yr} \times \dfrac{\$23.63}{h}$	$3,072
Supervisor	$0.15 \times \$3{,}070$	460.80
		—
Maintenance		
Maintenance Labor	$\dfrac{0.5\ h}{shift} \times \dfrac{shift}{8\ h} \times \dfrac{2{,}080\ h}{yr} \times \dfrac{\$24.25}{h}$	3,153
Materials	100% maintenance labor	3,153
Utilities		
Electricity	$\dfrac{7.050\ tons}{0.85} \times \dfrac{2.2\ kW}{ton} \times \dfrac{2{,}080\ h}{yr}$ $\times \dfrac{\$0.0701}{kWh}$	2,661
Total DAC		**$12,500**
Indirect Annual Costs, IAC		
Overhead	$0.6 \times (3{,}072 + 460.8 + 3{,}153 + 3{,}153)$	5,903
Administrative charges	$0.02\ (\$145{,}000)$	2,900
Property tax	$0.01\ (\$145{,}000)$	1,450
Insurance	$0.01\ (\$145{,}000)$	1,450
Capital recovery[a]	$0.0915 \times \$145{,}000$	13,270
Total IAC		**$24,970**
Recovery Credits, RC		
Recovered Acetone (rounded)	$\dfrac{300\ lb}{h} \times \dfrac{2{,}080\ h}{ye} \times \dfrac{\$0.26}{lb}$	($162,200)
Total Annual Cost (rounded)	$\$12{,}500 + \$27{,}620 - (\$162{,}200)$	**($122,000)** (Savings)

[a] For CRC, assumed interest rate is 4.25% and the equipment life is 15 years, for a capital recovery factor of 0.0915.

2.7　Example Problem 2

In this example problem, the alternate design procedure described in Section 2.3.8 is illustrated. The temperature of condensation is given, and the resultant removal efficiency is calculated. The example stream inlet parameters are identical to Example Problem 1 with the exception that removal efficiency is not specified and the required temperature of condensation is assumed to be 16 °F.

2.7.1　Required Information for Design

The first step is to calculate the partial pressure of the VOC at the specified temperature (16 °F) using Equation 2.7 to solve for P_{VOC}:

$$\log_{10} P_{VOC} = A - \frac{B}{T_{con} + C}$$

Remember to convert T_{con} to degrees Centigrade, i.e., 16 °F = -8.9 °C.

Substituting the values for the Antoine equation constants for acetone as listed in Table 2.6:

$$\log_{10} P_{VOC} = 7.11714 - \frac{1210.595}{-8.9 + 229.664}$$

$$P_{VOC} = 43.0 \; mm \; Hg$$

Using Equation 2.25, the removal efficiency is:

$$\eta = \frac{[760 \times (0.375)] - 43.0}{0.375 \times (760 - 43.0)} = 0.90$$

The remainder of the calculations in this problem are identical to those in Example Problem 1.

2.8　Acknowledgements

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Appendix A
Properties of Selected Compounds

Table A-1: Physical Properties for Selected Compounds

Compound	Critical Temp.[a] (°R)	Boiling Point (°F)	Molecular Weight (lb/lb-mole)	Heat of Condensation[b] (Btu/lb-mole)	Heat Capacity[c] (Btu/lb-mole-°F)	State
Acetone	915	134	58.08	12,520	30.19	Liquid
					17.81	Gas
Acetylene	555	-119	26.02	7,310	10.54	Gas
Acrylonitrile	965	171	53.06	14,030	15.25	Gas
Aniline	1258	364	93.13	18,240	45.87	Liquid
					25.79	Gas
Benzene	1012	176	78.11	13,210	32.51	Liquid
Benzonitrile	1259	375 [d]	103.12	19,750	26.08	Gas
Butane	765	31	58.12	9,640	23.31	Gas
Chloroethane	829	54	64.52	10,630	14.96	Gas
Chloroform	966	143	119.39	12,560	15.71	Gas
Chloromethane	749	-12	50.49	9,210	9.75	Gas
Cyclobutane	828	55	56.10	10,410	17.26	Gas
Cyclohexane	996	177	84.16	12,910	37.03	Liquid
					25.41	Gas
Cyclopentane	921	121	70.13	11,750	30.81	Liquid
					19.84	Gas
Cyclopropane	716	-27	42.08	8,650	13.29	Gas
Diethyl ether	840	94	74.12	11,400	41.26	Liquid
					28.57	Gas
Dimethylamine	787	44	45.09	11,360	16.90	Gas
Ethylbenzene	1111	277	106.17	15,320	30.69 [d]	Gas
Ethylene oxide	844	51	44.05	10,970	11.45	Gas
Heptane	972	209	100.12	13,680	53.76	Liquid
					39.68	Gas
Hexane	914	156	86.18	12,440	46.76	Liquid
					34.21	Gas
Methanol	923	148	32.04	15,150	19.41	Liquid
					10.54	Gas
Octane	1024	258	114.23	14,800	45.16	Gas
Pentane	845 [d]	97	72.15	11,100	28.73	Gas
Toluene	1065	231	92.14	14,290	37.53	Liquid
					24.76	Gas
o - Xylene	1135	292	106.17	15,880	44.49	Liquid
					31.86	Gas
m - Xylene	1111	282	106.17	15,360	43.82	Liquid
					30.50	Gas
p - Xylene	1109	281	106.17	15.360	30.33	Gas

[a] Reprinted with permission from the 16th edition of *Lange's Handbook of Chemistry*, Table 2.55 (Speight, 2005).
[b] Reprinted with permission from the 16th edition of *Lange's Handbook of Chemistry*, Table 2.54 (Speight, 2005). (Measured at boiling point).
[c] Reprinted with permission from the 16th edition of *Lange's Handbook of Chemistry*, Table 2.53 (Speight, 2005). (Measured at 77 °F).
[d] These values are taken from the 14th edition of *Lange's Handbook of Chemistry* (Dean, 1992).

Table A-2: Antoine Equation Constants for Selected Compounds[a]

Compound	Antoine Equation Constants			Valid Temperature Range (°F)
	A	B	C	
Acetone	7.11714	1210.595	229.664	Liquid
Acetylene	9.1402	1232.6	280.9	-202 to -117
	7.0999	711.0	253.4	-116 to -98
Acrylonitrile	7.03855	1232.53	222.47	-4 to 284
Aniline	7.32010	1731.515	206.049	216 to 365
Benzene	9.1064	1885.9	244.2	10 to 37
	6.90565	1211.033	220.790	46 to 217
Benzonitrile	6.74631	1436.72	181.0	Liquid
n-Butane	6.80898	935.86	238.73	-107 to 66
Chloroethane	6.98647	1030.01	238.61	-69 to 54
Chloroethylene	6.89117	905.01	239.48	-85 to 9
Chloroform	6.4934	929.44	196.03	-31 to 142
Chloromethane	7.09349	948.58	249.34	-103 to 23
Cyanic acid	7.56859	1251.86	243.79	-105 to 21
Cyclobutane	6.91631	1054.54	241.37	-76 to 54
Cyclohexane	6.84130	1201.53	222.65	68 to 178
Cyclopentane	6.88676	1124.162	231.36	-40 to 162
Cyclopropane	6.88788	856.01	246.50	-130 to -26
Diethyl ether	6.92032	1064.07	228.80	-78 to 68
Diethylamine	5.8016	583.30	144.1	88 to 142
Dimethylamine	7.08212	960.242	221.67	-98 to 44
1,4-Dioxane	7.43158	1554.68	240.34	68 to 221
Ethylbenzene	6.95719	1424.255	213.21	79 to 327
Ethylene oxide	7.12843	1054.54	237.76	-56 to 54
Heptane	6.89677	1264.90	216.54	28 to 255
Hexane	6.87601	1171.17	224.41	-13 to 198
Methanol	7.89750	1474.08	229.13	7 to 149
	7.97328	1515.14	232.85	147 to 230
Octane	6.91868	1351.99	209.15	66 to 306
Pentane	6.85296	1064.84	233.01	-58 to 136
Toluene	6.95464	1344.800	219.48	43 to 279
Vinyl acetate	7.2101	1296.13	226.66	72 to 162

Compound	Antoine Equation Constants			Valid Temperature Range (°F)
	A	**B**	**C**	
o-Xylene	6.99891	1474.679	213.69	90 to 342
m-Xylene	7.00908	1462.266	215.11	82 to 331
p-Xylene	6.99052	1453.430	215.31	81 to 331

[a] Reprinted with permission from the 16th edition of *Lange's Handbook of Chemistry*, Table 2.36 (Speight, 2005).

Appendix B
Documentation for Gasoline Vapor Recovery System Cost Data

As mentioned in Section 2.4.3, vendor cost data were obtained that related the equipment cost ($) of packaged gasoline vapor recovery systems to the process flow capacity (gal/ min). These data needed to be transformed, in order to develop Equation 2.35, which relates equipment cost ($) to system refrigeration capacity (R, tons), as follows:

$$EC_p = 4,910R + 212,000$$

To make this transformation, we needed to develop an expression relating flow capacity to refrigeration capacity. The first step was to determine the inlet partial pressure ($P_{VOC,in}$) of the VOC—gasoline, in this case. As was done in Section 3.1, we assumed that the VOC vapor was saturated and, thus, in equilibrium with the VOC liquid. This, in turn, meant that we could equate the partial pressure to the vapor pressure. The "model" gasoline had a Reid vapor pressure of 10 and a molecular weight of 66 lb/lb-mole, as shown in Section 7.1 of *Compilation of Air Pollutant Emission Factors* (EPA publication AP-42, Fifth Edition, November 2006). For this gasoline, the following Antoine equation constants were used:

$$A = 6.2236$$
$$B = 944.81$$
$$C = 233$$

These constants were obtained by extrapolating available vapor pressure vs. temperature data for gasoline found in Section 7.1 of AP-42. Upon substituting these constants and an assumed inlet temperature of 77 °F (25 °C) into the Antoine equation and solving for the inlet partial pressure ($P_{VOC,in}$). We obtain:

$$\log P_{VOC,in} = A - \frac{B}{T_{in} + C}$$

$$= 6.2236 - \frac{944.81}{25 + 233}$$

$$P_{VOC,in} = 365 \ mm \ Hg$$

If the system operates at atmospheric pressure (760 mm Hg), this partial pressure would correspond to a VOC volume fraction in the inlet stream of:

$$Y_{VOC,in} = \frac{365 \ mm}{760 \ mm} = 0.480$$

The outlet partial pressure ($P_{voc,out}$) and volume fraction are calculated in a similar way. The condensation (outlet) temperature used in these calculations is -80°F (-62°C), which is the typical operating temperature for the gasoline vapor recovery units for which the vendor supplied costs.

$$\log P_{VOC,out} = 6.2236 - \frac{944.81}{-62 + 233}$$

$$P_{VOC,out} = 5.01 \ mm \ Hg$$

This corresponds to a volume fraction in the outlet stream ($y_{VOC,out}$) of:

$$y_{VOC,out} = \frac{5.01\ mm}{760\ mm} = 0.0066$$

Substitution of $P_{VOC,out}$ and $y_{VOC,in}$ into Equation 2.25 yields the condenser removal efficiency:

$$\eta = \frac{(760 \times 0.480) - 5.01}{0.480\ (760 - 5.01)} = 0.993$$

The next step is determining the inlet and outlet VOC hourly molar flow rates ($M_{VOC,in}$ and $M_{VOC,out}$, respectively). As Equation 2.9 shows, $M_{VOC,in}$ is a function of $y_{VOC,in}$ and the total inlet volumetric flow rate, Q_{in}, (scfm).

Now, because the gasoline vapor flow rates are typically expressed in gallons/minute, we have to convert them to scfm. This is done as follows:

$$Q_{in} = Q_g \left(\frac{gal}{min} \right) \times \frac{1\ ft^3}{7.48\ gal} = 0.134\ Q_g\ scfm$$

Substituting these variables into Equation 2.9, we obtain:

$$M_{VOC,in} = \frac{0.134\ Q_g}{392} (0.480)(60) = 0.00984\ Q_g \left| \frac{lb - mole}{hr} \right|$$

We obtain $M_{VOC,out}$ from Equation 2.2:

$$M_{VOC,out} = 0.00984\ Q_g (1 - 0.993) = 6.89 \times 10^{-5} Q_g \left(\frac{lb - moles}{hr} \right)$$

And according to Equation 2.10, the amount of gasoline vapor condensed ($M_{VOC,recovered}$) is the difference between $M_{VOC,in}$ and $M_{VOC,out}$:

$$M_{VOC,recovered} = 0.00977\ Q_g \left(\frac{lb - moles}{hr} \right)$$

The final step is to calculate the condenser heat load. This load is a function of the inlet, outlet, and condensate molar flow rates, the inlet and condensation temperatures, the heat capacities of the VOC and air, and the VOC heat of condensation. The VOC heat capacity used are based on pentane and butane chemical properties, the largest components of gasoline, and were obtained from *Lange's Handbook of Chemistry* (16th edition). The heat of condensation data were obtained for gasoline from *CHRIS Hazardous Chemical Data* (U.S. Coast Guard, U.S. Department of Transportation, June 1999).

Heat capacities (Btu/lb-mole-°F):[5]

$$C_{p,VOC} = 26.6$$
$$C_{p,air} = 6.95$$

Heat of condensation of VOC (Btu/lb-mole):[6]

$$\Delta H_{VOC} = 9,240$$

Substitution of these data, the molar flow rates, and the temperatures into Equations 2.12, 2.16 and 2.17 yields the following enthalpy changes in Btu/hr:

$$\Delta H_{con} = 131.1\, Q_g$$
$$\Delta H_{uncon} = 0.288\, Q_g$$
$$\Delta H_{noncon} = 11.6\, Q_g$$

The condenser heat load (H_{load}) is the sum of these three enthalpy changes (Equation 2.11):

$$H_{load} = 143\, Q_g$$

The refrigeration capacity (R, tons) is computed from Equation 2.23:

$$R = \frac{H_{load}}{12,000} = 0.0119\, Q_g$$

This last equation relates the refrigeration capacity (tons) to the inlet gasoline vapor flow rate (gal/min). Solving for Q_g, in terms of R, we obtain:

$$Q_g = 83.9R$$

Finally, we substitute this relationship into the equipment cost (\$) vs. vapor flow rate (Q_g) correlation, which was developed from the vendor cost data:

$$EC_p = 58.5Q_g + 212,000$$
$$= 58.5(83.9R) + 212,000$$
$$= 4,910\,R + 212,000$$

Note that this last expression is identical to Equation 2.35 (and that Equation 2.35 has been scaled to 2014 dollars).

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[5] The heat capacity for VOC given here is based on C_p for butane of 23.29 Btu/lb-mole-°F and C_p for pentane of 28.73 Btu/lb-mole-°F, giving an average C_p for gasoline of 26.6 Btu/lb-mole-°F. See Table A-1 in Appendix A. *Lange's Handbook of Chemistry* (16th edition).

[6] For "Gasoline: Automotive (<4.23 grams lead/gallon)," the Latent Heat of Vaporization is given as a range of 130 to 150 Btu/lb. Assuming the MW of gasoline vapor is 66 lb/lb-mole and the average heat of vaporization is 140 Btu/lb, the result is 9,240 Btu/lb-mole. *CHRIS Hazardous Chemical Data* (U.S. Coast Guard, U.S. Department of Transportation, 1999).