

Heat Exchangers and Energy Management

5.1 Introduction

Step 3 of our plantwide control design procedure involves two activities. The first is to design the control loops for the removal of heat from exothermic chemical reactors. We dealt with this problem in Chap. 4, where we showed various methods to remove heat from exothermic reactors and how to control the temperature in such reactors. At that point we assumed that the heat was removed directly and permanently from the process (e.g., by cooling water). However, it is wasteful to discard the reactor heat to plant utilities when we need to add heat in other unit operations within the process. Instead, a more efficient alternative is to heat-integrate various parts of the plant by the use of process-to-process heat exchangers.

The second activity in Step 3 therefore deals with the management of energy in heat-integrated processes. The reason we look at energy management in conjunction with heat removal controls for exothermic reactors is that these issues are often closely coupled. For example, it is common to preheat the feed to adiabatic tubular reactors with the reactor effluent stream. This recovers heat from the reactor that otherwise would be discarded to utilities. However, it also retains the heat within the process and could cause excessive temperatures around the reactor unless properly managed. Another potential problem is the thermal feedback provided by the feed-effluent exchanger. As we saw in Chap. 4, such feedback can cause multiplicity and open-loop instability of the reactor. The reactor control system must therefore be revisited in light of the heat integration schemes considered.

Heat management is accomplished by controlling the flow of energy

in various heat exchangers. We differentiate between utility exchangers and process-to-process heat exchangers. Utility exchangers allow us to import heat from heat sources or discard heat to the environment. Process-to-process exchangers, on the other hand, are used to transfer heat from one part of the process to another. We will look at the unit operation controls of both types of exchangers.

We will not discuss the steady-state design techniques involved in arriving at a heat-integrated process. This is adequately covered in texts on process design (e.g., Douglas, 1988). We will, however, examine the dynamic and control implications of such integrated designs. In particular, we will point out design modifications required to make a heat-integrated plant operable and controllable.

5.2 Fundamentals of Heat Exchangers

Process understanding precedes control system design. The area of heat management is no exception. To provide the proper background for this chapter we therefore start with a brief review of heat exchanger characteristics and the thermodynamic foundations of heat management. We examine the thermal efficiency of a typical chemical process and explore steady-state incentives for heat conservation and integration. We also extend the discussion from Chap. 4 regarding the role of thermodynamics in process control.

5.2.1 Steady-state characteristics

Figure 5.1 shows a typical shell-and-tube exchanger used to exchange sensible heat between two streams. The familiar design equations are

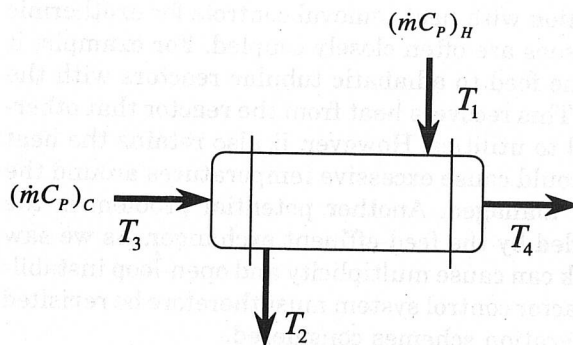


Figure 5.1 Schematic of a shell-and-tube heat exchanger.

$$Q = (\dot{m}C_p)_H(T_1 - T_2) = (\dot{m}C_p)_C(T_4 - T_3) \quad (5.1)$$

$$= UA \frac{(T_1 - T_4) - (T_2 - T_3)}{\ln \frac{T_1 - T_4}{T_2 - T_3}} \quad (5.2)$$

where Q = heat transferred
 T_1, T_2, T_3, T_4 = stream temperatures
 $(\dot{m}C_p)_H$ = product of flowrate and specific heat capacity for hot stream
 $(\dot{m}C_p)_C$ = product of flowrate and specific heat capacity for cold stream

Equations (5.1) and (5.2) are ideally suited to design the exchanger and thereby calculate the heat transfer area required to meet the specified stream temperatures. However, the equations are less convenient as rating equations to determine the steady-state exit temperatures for a given exchanger under different operating conditions. This is of course what we need in control studies when we must determine the process gain and the steady-state effects of disturbances. For those applications it is easier to use the effectiveness method (Gebhart, 1971; Jones and Wilson, 1997). The effectiveness of an exchanger is defined as

$$\epsilon = \frac{(\dot{m}C_p)_H(T_1 - T_2)}{(\dot{m}C_p)_{\min}(T_1 - T_3)} = \frac{(\dot{m}C_p)_C(T_4 - T_3)}{(\dot{m}C_p)_{\min}(T_1 - T_3)} \quad (5.3)$$

where $(\dot{m}C_p)_{\min}$ is the smaller of $(\dot{m}C_p)_H$ and $(\dot{m}C_p)_C$.

For given values of the inlet flows and temperatures, the exit temperatures are explicitly calculated for a known exchanger effectiveness:

$$T_2 = T_1 - \frac{\epsilon(\dot{m}C_p)_{\min}(T_1 - T_3)}{(\dot{m}C_p)_H} \quad (5.4)$$

$$T_4 = T_3 + \frac{\epsilon(\dot{m}C_p)_{\min}(T_1 - T_3)}{(\dot{m}C_p)_C} \quad (5.5)$$

It can be shown that the effectiveness is determined by the exchanger's design parameters through the following equation:

$$\epsilon = \frac{1 - e^{-(1-r)N_{TU}}}{1 - re^{-(1-r)N_{TU}}} \quad (5.6)$$

where $r = \frac{(\dot{m}C_p)_{\min}}{(\dot{m}C_p)_{\max}}$

$$N_{TU} = \frac{UA}{(\dot{m}C_p)_{\min}}$$

When the two streams exchanging heat have the same mass flowrate and heat capacity, $r = 1$, and the effectiveness is calculated from

$$\epsilon = \frac{UA/\dot{m}C_p}{1 + UA/\dot{m}C_p} \quad (5.7)$$

5.2.2 Heat exchanger dynamics

Heat exchangers have fast dynamics compared to other unit operations in a process. Normally the time constant is measured in seconds but could be up to a few minutes for large exchangers. Process-to-process exchangers should be modeled rigorously by partial differential equations since they are distributed systems. This introduces the correct amount of deadtime and time constant in the exit stream temperatures, but the models are inconvenient to solve. We have found that for the purpose of plantwide control studies it is not necessary to have such detailed descriptions of the exchanger dynamics, since these units rarely dominate the process response. Instead, it is often possible to construct useful models by letting two sets of well-stirred tanks in series exchange heat. This simplifies the solution procedure. Alternatively, we can use the effectiveness method to calculate the steady-state exchanger exit temperatures and then delay these temperatures by first-order time constants to capture the dynamics.

5.3 Thermodynamic Foundations

5.3.1 Energy, work, and heat

There are two aspects of energy that we shall be interested in, heat and work. Work is the hard currency of energy and it can always be completely converted into heat. Heat, on the other hand, is a less organized source of energy and it can only partly be converted into work. The remaining heat has to be discarded into a low-temperature heat sink. The largest fraction of work that can be extracted from a heat unit depends upon the absolute temperatures of the heat source and the heat sink.

$$\left(\frac{w}{q}\right)_{\max} = \frac{T_1 - T_2}{T_1} \quad (5.8)$$

where w = work
 q = heat
 T_1 = absolute temperature of heat source
 T_2 = absolute temperature of heat sink

the same mass flowrate is calculated from

(5.7)

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Since the work portion of energy is independent of the method of delivering the energy, it is most meaningful to compare the energy usage of various unit operations on a work-equivalent basis. We can then differentiate between work producers and work consumers. As a matter of generalization, we can state that it is possible, in principle, to extract work from spontaneous processes and that work is always needed to reverse such processes. When we cannot extract work from a spontaneous process, the work potential is lost.

What, then, are the spontaneous processes in a chemical plant? Here is a short but complete list.

- Chemical reactions
- Mixing of streams with different compositions, temperatures, and pressures
- Flow of material from high to low pressure
- Flow of heat from high to low temperature

We are rarely able to extract much work from chemical reactors. We mostly take out energy from the reactor in the form of heat. Also, when we let a hot stream heat up a cold stream in a heat exchanger we lose some of the work potential of the hot stream. The same is true for mixing and material flow across a pressure drop; we take advantage of the spontaneity of the process and make no attempt to recover work from it. In contrast, we often want to perform operations that are the reverse to the spontaneous direction. This always requires work. For example, separation is the opposite of mixing. The work demand of separating an ideal mixture of n components into pure products at constant temperature T is

$$w = -RT \sum_{i=1}^n x_i \ln x_i \quad (5.9)$$

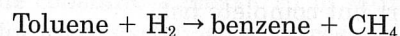
where w = separation work per mole of mixture
 R = universal gas-law constant
 T = absolute temperature of the separation
 x_i = mole fraction of component i in the mixture

When we consider that we lose work potential in chemical reactors, in heat exchangers, and in mixing operations and combine this with the need for work in separation processes, pumping, and compression, it becomes clear that chemical processes are usually very inefficient from an energy standpoint. Of course, energy conservation is usually of secondary importance in chemical processing; safety, quality, and productivity are more important. Nevertheless, it is economically sound

to try to improve the thermal efficiency of the plant. We just have to do it in a way that does not jeopardize the primary goals of the operation.

5.3.2 HDA example

To appreciate better the use of energy in a chemical plant and the opportunities that exist for heat integration, we will look at a specific case. The HDA process introduced in Chap. 1 will serve as our example. We recall the main reaction in this process is



$$\text{Standard state heat of reaction } \Delta H^0 = -42.2 \text{ kJ/mol}$$

$$\text{Change in Gibbs free energy } \Delta G^0 = -43.1 \text{ kJ/mol}$$

As mentioned in Chap. 4, ΔH^0 is the amount of heat that has to be removed to operate a reactor isothermally at the standard states of the components. The Gibbs free energy of reaction, ΔG^0 , on the other hand, is how much work we could extract in the form of electrical energy from a hypothetical reactor operating as a reversible, isothermal fuel cell. Since we don't know how to build fuel cells around arbitrary reactions we usually extract the reaction energy as heat. Some of this heat can, of course, be converted to mechanical or electrical energy by use of a turbine and a generator. However, we take a healthy cut in our dividends. For example, let's say we made 500 psia steam from the hot effluents of the HDA reactor. Equation (5.8) gives us the maximum turbine work we could extract from the steam.

$$w \approx 42.2 \cdot \frac{515 - 298}{515} = 17.8 \text{ kJ/mol toluene}$$

Notice that we have used the standard state heat of reaction to estimate the heat evolved at 515 K. This is only approximately true since the standard state heat of reaction should be adjusted for the difference in heat capacities between the products and the reactants at the high temperature. The only point we are trying to convey is that the available work is significantly reduced when we use steam to drive a turbine. Furthermore, because of inefficiencies of the turbine and the generator, the electrical energy generated this way ends up as a small fraction of the 43.1 kJ/mol toluene that is theoretically available from the reaction.

We next turn to the separation section of the process. Equation (5.9) can help us get a rough estimate of what the separation costs would be. We first look at the total flow exiting the reactor and consider how much work would be required to separate this stream into pure

components. Next, we consider the work associated with the vapor recycle stream. This work is subtracted from the total work of the exit stream since we don't have to separate the components in the gas recycle. When we use the flowrate and composition data from Chap. 10 and we assume that the separation is being done at a constant temperature (298 K), we get the following results by using Eq. (5.9):

$$\begin{aligned}
 w_{\text{Rx Exit}} &= -F \cdot R \cdot T \cdot \sum_{i=1}^5 x_i \ln x_i \\
 &= -553 \text{ mol/s} \cdot 8.31 \text{ J/mol K} \cdot 298 \text{ K} \\
 &\quad \cdot [0.364 \ln (0.364) + 0.546 \ln (0.546) + 0.069 \ln (0.069) \\
 &\quad + 0.019 \ln (0.019) + 0.002 \ln (0.002)] \\
 &= 553 \cdot 8.31 \cdot 298 \cdot 0.97 = 1.3 \times 10^6 \text{ J/s} \\
 w_{\text{Recycle}} &= 444 \cdot 8.31 \cdot 298 \cdot 0.71 = 0.78 \times 10^6 \text{ J/s} \\
 w_{\text{Separation}} &= (1.3 - 0.78) \times 10^6 = 0.52 \text{ MW}
 \end{aligned}$$

The net work of 0.52 MW for the separation section assumes that we perform all separations mechanically, i.e., with compressors and semipermeable membranes. In reality we use evaporation, partial condensation, and distillation to separate the components. We can use Eq. (5.8) to estimate how much heat, from 75 psia steam, we must supply to an ideal separation device to provide the necessary separation work. Again, this is only a rough estimate of the required heat.

$$q \approx w \cdot \frac{T_1}{T_1 - T_0} = 0.52 \cdot \frac{426}{426 - 298} = 1.7 \text{ MW}$$

In reality, partial condensers and distillation columns are far from ideal separation devices. There is plenty of mixing and heat degradation in these operations to lower the efficiency significantly. In fact, we can see from Table 10.6 in Chap. 10 that the net heat consumption for partial condensation and distillation is over 5 MW.

We can summarize the HDA example as follows. The process converts 132 kmol/h of toluene. If it were possible to extract all the work contained in the reactants we could generate 1.58 MW of electric power at standard conditions. When forced to generate power from steam we would obtain much less than 0.65 MW of electric power. When none of the reaction energy is converted to work we need to dissipate about 1.55 MW of heat to utilities. On the separation side we need 0.52 MW of mechanical power at standard conditions. If ideal heat-driven separation devices were

available they would consume roughly 1.7 MW of heat. In actual distillation columns the separation cost is over 5 MW of heat.

This example shows clearly why most chemical processes are so thermally inefficient. In theory we have an excess of work available ($1.58 - 0.52 > + 1$ MW). In reality we have a deficiency in heat ($1.55 - 5 < - 3.45$ MW).

So far we have looked at the energy requirements of the process operating at ambient conditions (i.e., 25°C, 1 atm). However, these conditions are impractical other than for raw material and product storage. For example, the rate of reaction between toluene and hydrogen is not appreciable until the reactor inlet conditions are around 600°C and over 30 atm. However, this high temperature is impractical for material storage and for distillation. Distillation columns are mostly operated in the 50 to 200°C range since steam is commonly used in the reboilers and cooling water in the condensers. The bottom line is that, for practical reasons, we need to operate various processing steps at different conditions. The transition of process streams between these different conditions costs energy. To visualize just how much we need for this purpose we refer to the temperature-enthalpy diagram (*T-H* diagram) for the HDA process (Terrill and Douglas, 1987a; Douglas, 1988). A *T-H* diagram shows how much heat is required to bring the cold streams in the process to higher temperatures and how much cooling is needed to bring the hot streams down to lower temperatures. We find that as much as 26 MW of heating and cooling are required to transition streams within the HDA plant!

5.3.3 Heat pathways

It is important to realize that there are no thermodynamic restrictions on the energy requirement to transition streams between unit operations. In other words, the heating and cooling of streams is done for practical reasons and not to satisfy the laws of thermodynamics. This energy would not be an issue if all the processing steps operated at the same constant temperature. Furthermore, since raw materials and products are stored at roughly the same temperature, the net energy requirement for heating and cooling equals the heat losses from the process. We therefore realize that most of the energy required for heating certain streams within the process is matched by a similar amount required for cooling other streams. Heat recovered from cooling a stream could be recycled back into the process and used to heat another stream. This is the purpose of heat integration and heat exchanger networks (HENs).

From a plantwide perspective we can now discern three different

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Figure 5.2

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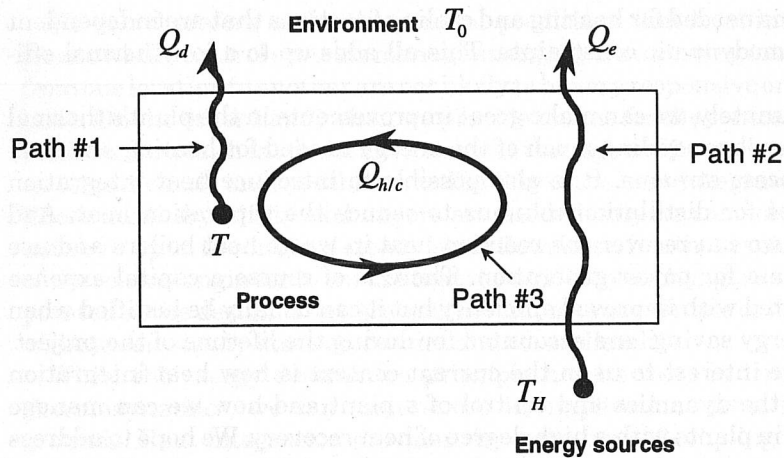


Figure 5.2 Heat pathways.

“heat pathways” in the process. See Fig. 5.2 for an illustration. The first pathway dissipates to the environment heat generated by exothermic reactions and by degradation of mechanical work (e.g., compression, pressure drop, and friction). This pathway is from inside the process and flows out. It is of course possible to convert some of the heat to work as it is removed from high temperatures in the process.

A second pathway carries heat from utilities into the process. Mechanical work is extracted from the heat as it flows from a high supply temperature to the lower temperature of the environment. This pathway goes through the process and is needed to satisfy the thermodynamic work requirements of separation. Work is also extracted from the heat stream to overcome process inefficiencies associated with stream mixing and heat transfer.

The third pathway is internal to the process. Here heat flows back and forth between different unit operations. The magnitude of this energy path depends upon the heating and cooling needs and the amount of heat integration implemented. Whenever the internal path is missing, and there is a heating requirement, the heat has to be supplied from utilities. The same amount of heat must eventually be rejected to the environment elsewhere in the process.

5.3.4 Heat recovery

We have tried to illustrate why chemical processes are thermally inefficient. First, the chemical work available in the reactants is dissipated as heat. Second, the work required for separation is usually supplied as heat to distillation columns, which have internal inefficiencies. Finally,

energy is needed for heating and cooling functions that are independent of thermodynamic constraints. This all adds up to a low thermal efficiency.

Fortunately, we can make great improvements in the plant's thermal efficiency by recycling much of the energy needed for heating and cooling process streams. It is also possible to introduce heat integration schemes for distillation columns to reduce the separation heat. And finally we can recover the reaction heat in waste heat boilers and use the steam for power generation. There is of course a capital expense associated with improved efficiency but it can usually be justified when the energy savings are accounted for during the lifetime of the project. Of more interest to us in the current context is how heat integration affects the dynamics and control of a plant and how we can manage energy in plants with a high degree of heat recovery. We hope to address these concerns adequately in the remainder of this chapter. But before we embark on heat exchanger and energy management control methods we extend the discussion from Chap. 4 regarding thermodynamics and process control.

5.3.5 Exergy destruction principle

In Chap. 4 we introduced the concept of exergy in connection with the design of responsive control loops. Exergy is the mechanical work potential of a stream or the energy stored in a process relative to the conditions in the earth's environment. It incorporates all the work functions we have looked at so far, that is, work from heat [Eq. (5.8)], separation work [Eq. (5.9)], and the Gibbs free energy change of reaction. As opposed to plain "work," exergy is a thermodynamic state function (see Appendix A for more details on this). The reason exergy is a state function, and work is not, comes from the fact that exergy expresses the work potential of a stream *relative* to the equilibrium state of the environment. Without such a reference, the work portion of energy is undetermined.

Exergy and energy are expressed in the same units but the two entities are different in one important aspect. Energy is a conserved property, exergy is not. In fact, all spontaneous processes are accompanied by a degree of exergy destruction. While this property is important in identifying the thermal efficiencies of a process, it also has an interesting implication for process control. We showed in Chap. 4, and discuss further in Appendix A, that responsive control is achieved when the control system can significantly *alter the rate of exergy destruction* in the process. A limited amount of exergy is destroyed when energy is kept within the process by means of heat transfer between hot and cold streams. A maximum amount of exergy is destroyed when heat is dissipated from the process to the environment.

We can therefore make the general prediction that heat-integrated plants whose control systems manage energy by transferring excess heat from one location to another are not likely to be very responsive or controllable. On the other hand, when we provide means for the control system to discard excess heat to the environment through utility exchangers we can significantly improve the controllability of the heat-integrated plant. These auxiliary utility exchangers are needed only for control and will not show up on the original steady-state flowsheet. It is the responsibility of the control engineer to ensure that the auxiliary exchangers are incorporated in the design. They are another important example of the trade-off between steady-state economics and controllability.

While the exergy destruction principle has its roots in irreversible thermodynamics, we can understand its implications from intuitive grounds. Heat integration provides the same type of coupling between units that occurs with material recycle. A disturbance in one unit is recycled to another unit which in turn can affect the first unit. In material recycle loops we limit the effects of flow disturbances by flow controlling somewhere in the loop. Composition disturbances are harder to deflect and here we mostly rely on the unit operation controls and the plantwide control strategy to minimize their propagation. Energy disturbances will also be recycled unless they can be deflected by the control system. Fortunately there is a simple way of doing this. We can sense the disturbance, typically as a change in temperature somewhere in the loop, and divert the disturbance to the utility system by way of utility exchangers.

5.4 Control of Utility Exchangers

Armed with the thermodynamic fundamentals of heat management, we now take a closer look at the unit operation control loops for heat exchangers. We start with utility exchangers. These are used when heat is supplied to, or removed from, the process. Examples are steam-heated reboilers, electric heaters, fuel-fired furnaces, water-cooled condensers, and refrigerated coolers.

The purpose of unit operation controls is to regulate the amount of energy supplied or removed. This is typically done by measuring a temperature in the process and manipulating the flowrate of the utility. A PI-controller is adequate in most instances although derivative action can be used to compensate for the lag introduced by the thermowell. The location of the temperature measurement depends upon the purpose of the heat exchange. For example, when reaction heat is dissipated through a water cooler, the controlled temperature is in the reactor. We gave many examples of this in Chap. 4. Similarly, when the utility exchanger delivers heat to a separation system, the control point should be

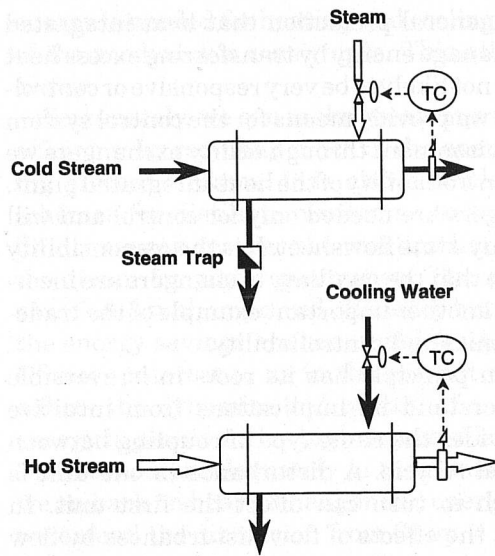


Figure 5.3 Control of utility exchangers.

located where the effects of the added energy are felt the most. Control of a tray temperature in a distillation column by manipulation of reboiler steam is a good example. Finally, when the utility exchanger is used for stream heating and cooling, the control point is on the stream being heated or cooled. Figure 5.3 shows two examples of the latter application.

In some situations the heat load on a utility exchanger is controlled without adjusting the utility flow. Consider the partial condenser on a distillation column shown in Fig. 5.4. The cooling water valve is run

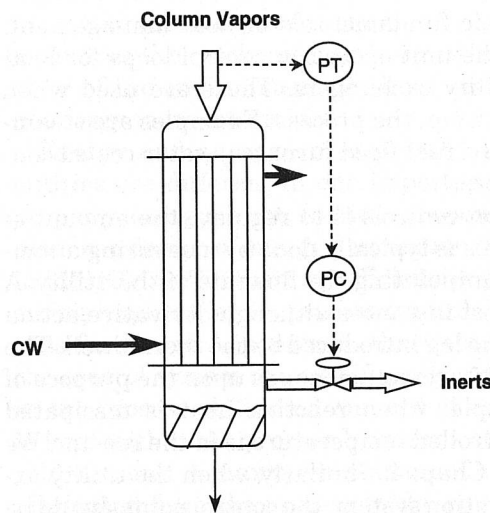


Figure 5.4 Indirect control of heat transfer rate in partial condenser with inerts.

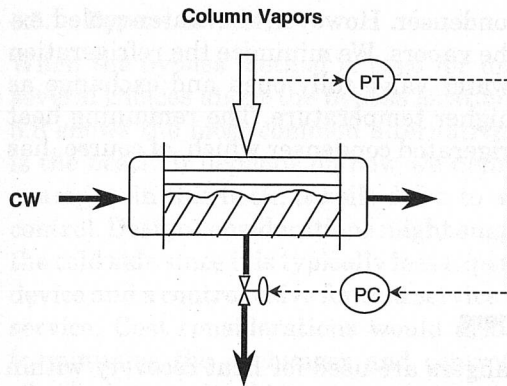


Figure 5.5 Heat transfer control by variable heat transfer area in flooded condenser.

wide open and heat transfer is affected by adjusting the inert concentration in the condenser. This arrangement ensures the best recovery of condensables in the presence of inerts. When there are no inerts present in the column, heat transfer control can be done by flooding the condenser as in Fig. 5.5. By manipulating the condensate flow, the heat transfer area is adjusted to provide the desired amount of heat transfer for condensation at a constant cooling water flow. However, the final temperature of the condensate (the degree of subcooling) will vary with the condensing load.

Occasionally utility exchangers need no control loops at all. For example, the distillation column in Fig. 5.6 uses an uncontrolled cooling water condenser. The water-cooled condenser reduces the expensive

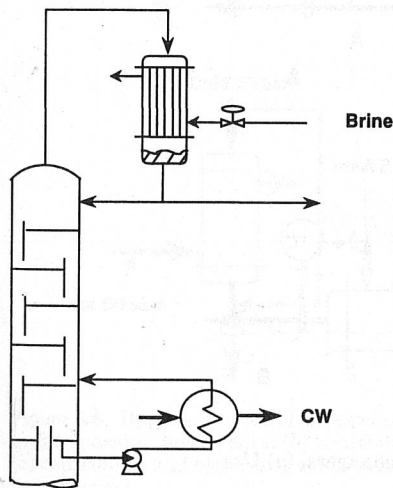


Figure 5.6 Utility exchanger without heat transfer control.

duty of the top refrigerated condenser. However, the water-cooled exchanger cannot condense all the vapors. We minimize the refrigeration load by keeping the cooling water valve fully open and exchange as much heat as we can at the higher temperature. The remaining heat has to be taken out by the refrigerated condenser which, of course, has to be controlled somehow.

5.5 Control of Process-to-Process Exchangers

Process-to-process (P/P) exchangers are used for heat recovery within a process. Equations (5.4) and (5.5) tell us that we can control the two exit temperatures provided we can independently manipulate the two inlet flowrates. However, these flowrates are normally unavailable for us to manipulate and we therefore give up two degrees of freedom for temperature control compared to the case where we use utility exchangers on both streams. We can restore one of these degrees of freedom fairly easily. For example, it is possible to oversize the P/P exchanger and provide a controlled bypass around it as in Fig. 5.7a. It is also possible to combine the P/P exchanger with a utility exchanger as in Fig. 5.7b.

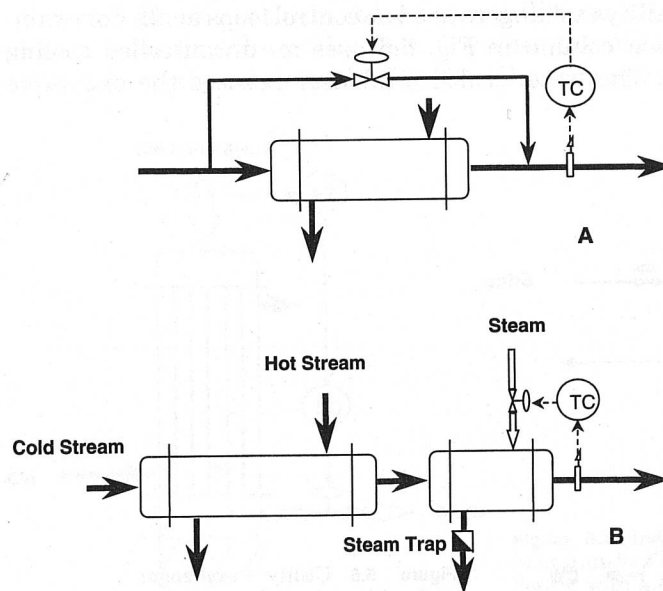


Figure 5.7 Control of process-to-process heat exchangers. (a) Use of bypass; (b) use of auxiliary utility exchanger.

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5.5.1 Bypass control

When the bypass method is used for unit operation control, we have several choices about the bypass location and the control point. Figure 5.8 shows the most common alternatives. We may ask "Which option is the best?" It depends on how we define "best." As with many other examples in this book, it boils down to a trade-off between design and control. Design considerations might suggest we measure and bypass on the cold side since it is typically less expensive to install a measurement device and a control valve for cold service than it is for high-temperature service. Cost considerations would also suggest a small bypass flow to minimize the exchanger and control valve sizes. From a control standpoint we should measure the most important stream, regardless of temperature, and bypass on the same side as we control (e.g., Fig. 5.8a and c). This minimizes the effects of exchanger dynamics in the loop. We should also want to bypass a large fraction of the controlled stream since it improves the control range. This requires a large heat exchanger.

There are several general heuristic guidelines for heat exchanger bypass systems. However, this very much remains an open research area since these guidelines are not always adequate to deal with all of

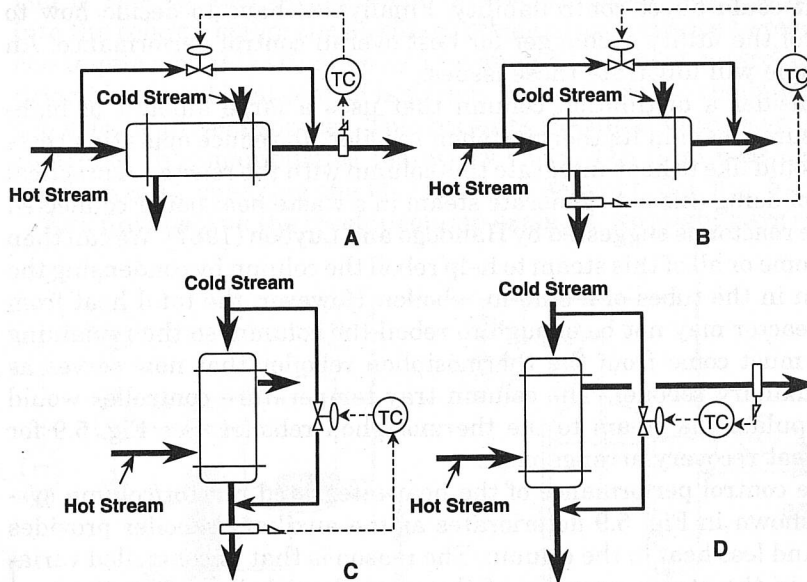


Figure 5.8 Bypass control of process-to-process heat exchangers. (a) Controlling and bypassing hot stream; (b) controlling cold stream and bypassing hot stream; (c) controlling and bypassing cold stream; (d) controlling hot stream and bypassing cold stream.

the issues for bypass systems. We typically want to bypass the flow of the stream whose temperature we want to control. The bypass should be about 5 to 10 percent of the flow to be able to handle disturbances. Finally, we must carefully consider the fluid mechanics of the bypass design for the pressure drops through the control valves and heat exchanger. Instead of using just one valve in the bypass line, we also can consider two valves or a three-way valve to provide sufficient rangeability of flows (Jones and Wilson, 1997).

5.5.2 Use of auxiliary utility exchangers

When the P/P exchanger is combined with a utility exchanger, we also have a few design decisions to make. We must first establish the relative sizes between the recovery and the utility exchangers. From a design standpoint we would like to make the recovery exchanger large and the utility exchanger small. This gives us the most heat recovery, and it is also the least expensive alternative from an investment standpoint. However, a narrow control range and the inability to reject disturbances make this choice the least desirable from a control standpoint.

Next, we must decide how to combine the utility exchanger with the P/P exchanger. This could be done either in a series or parallel arrangement. Physical implementation issues may dictate this choice but it could affect controllability. Finally, we have to decide how to control the utility exchanger for best overall control performance. An example will illustrate these issues.

Consider a distillation column that uses a large amount of high-pressure steam in its thermosiphon reboiler. To reduce operating costs we would like to heat-integrate this column with the reactor. A practical way of doing this is to generate steam in a waste heat boiler connected to the reactor as suggested by Handogo and Luyben (1987). We can then use some or all of this steam to help reboil the column by condensing the steam in the tubes of a stab-in reboiler. However, the total heat from the reactor may not be enough to reboil the column, so the remaining heat must come from the thermosiphon reboiler that now serves as an auxiliary reboiler. The column tray temperature controller would manipulate the steam to the thermosiphon reboiler. See Fig. 5.9 for the heat recovery arrangement.

The control performance of the heat-integrated reactor/column system shown in Fig. 5.9 deteriorates as the auxiliary reboiler provides less and less heat to the column. The reason is that uncontrolled variations in the steam pressure of the waste heat boiler affect the heat supplied to the column. When these variations are of the same order of magnitude as the total heat supplied by the auxiliary reboiler, the latter cannot compensate properly for the variations. Part of the prob-

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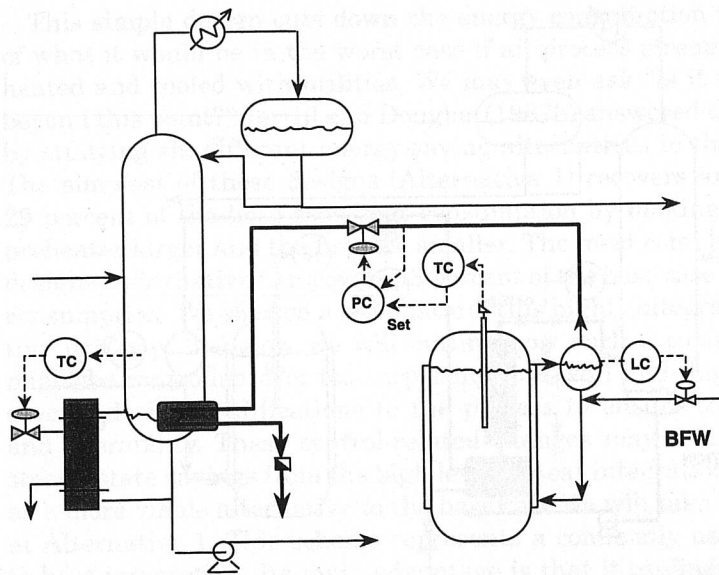


Figure 5.9 Reactor/column heat integration with auxiliary reboiler in parallel.

blem in this parallel arrangement is that the disturbances propagate into the column before the auxiliary reboiler has a chance to react. We can improve on this situation by providing a total heat controller. The principle of total heat control is simple. We measure the combined heat input from all heat sources. This becomes the input to the so-called Q controller that manipulates the utility valve. The total heat demand is adjusted by changing the Q controller's setpoint. See Fig. 5.10.

As a final note on the reactor/column example, we might have consid-

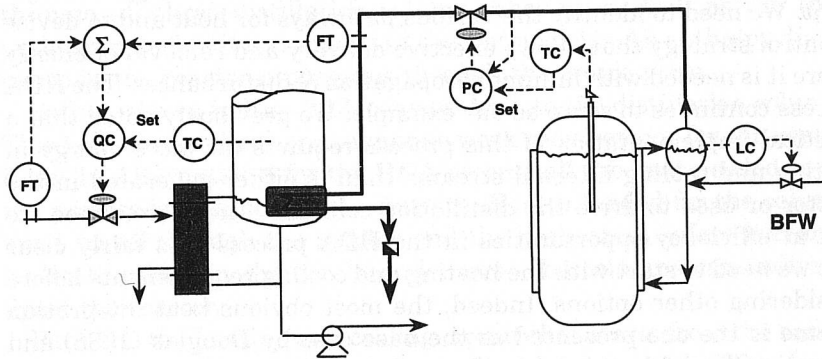


Figure 5.10 Reactor/column heat integration with auxiliary reboiler in parallel and Q controller.

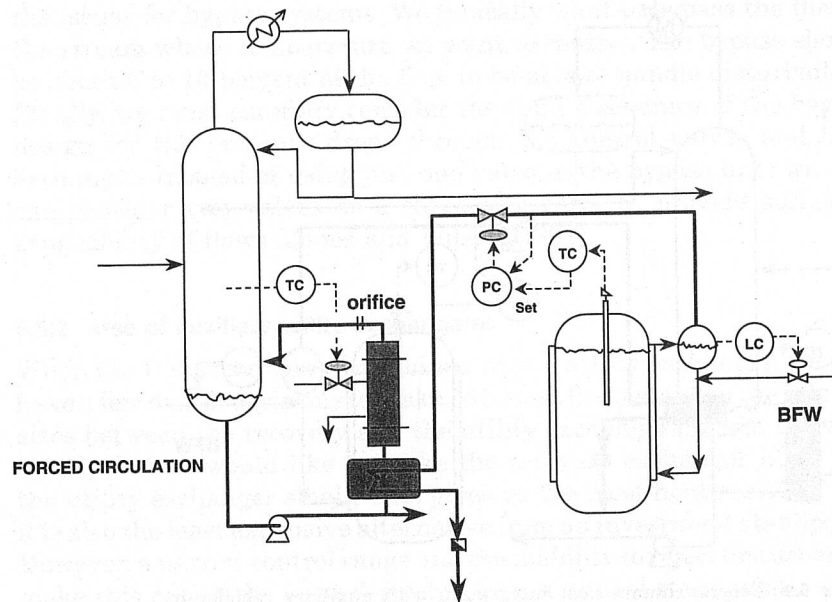


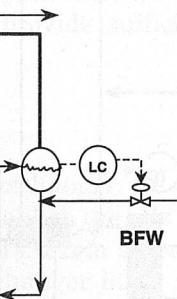
Figure 5.11 Reactor/column heat integration with auxiliary reboiler in series.

ered a series arrangement as in Fig. 5.11. Temperature variations from the reactor would now be attenuated in the auxiliary heater by virtue of changes in the temperature driving force for heat transfer.

5.6 Plantwide Energy Management

5.6.1 Introduction

We now look at the plantwide control issues around energy management. We need to identify the various pathways for heat and to devise a control strategy that allows effective delivery and removal of energy where it is needed with minimal propagation of disturbances. The HDA process continues to serve as our example. We previously noted that a practical implementation of this process requires far more energy in heating and cooling internal streams than is either generated in the reactor or used to drive the distillation columns. Therefore, when we look at efficiency opportunities in the HDA process it is fairly clear that we need to start with the heating and cooling requirements before considering other options. Indeed, the most obvious heat integration scheme is the one presented as the base case by Douglas (1988) and shown in Chap. 1 in this book. Here we use the hot effluent gases from the reactor to help preheat the recycle streams and the fresh feeds before they enter the reactor.



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This simple design cuts down the energy consumption to one-third of what it would be in the worst case if all process streams had to be heated and cooled with utilities. We may even ask "Is it worth going beyond this point?" Terrill and Douglas (1987b) answered this question by studying six different energy-saving alternatives to the base case. The simplest of these designs (Alternative 1) recovers an additional 29 percent of the base case heat consumption by making the reactor preheater larger and the furnace smaller. The most complicated of the designs (Alternative 6) recovers 43 percent of the base case's net energy consumption. We showed a schematic of this highly integrated alternative in Chap. 2. Below we will outline how such a complex scheme might be controlled. Not too surprisingly we will find that it requires several design modifications to the process to ensure controllability and operability. These control-related changes may easily offset the steady-state savings from the high level of heat integration. Therefore, as a more viable alternative to the base case we will take a closer look at Alternative 1. This scheme represents a commonly used approach to heat integration. Its main advantage is that it confines the recycle of heat to the reactor. The disadvantage is that it provides a high degree of thermal feedback to the reactor which can be detrimental to the reactor's behavior (e.g., output multiplicity and open-loop instability).

5.6.2 Controlling plantwide heat integration schemes

In this section we outline the approach we would take in controlling a complex heat-integrated scheme like Alternative 6 of the HDA process. The first step is to identify the three heat pathways.

Path 1 is intended to carry heat from exothermic reactions that must be dissipated from the process. This path, shown in Fig. 5.12, goes through all three distillation column reboilers as well as the three preheaters before it terminates in the water cooler. An enthalpy disturbance in the reactor propagates through the entire plant.

Path 2, shown in Fig. 5.13, conveys heat to the distillation columns. This heat covers the thermodynamic work requirement for the separations. In Alternative 6 for the HDA process all the heat needed to run the process is supplied by the furnace. Heat intended for the columns must travel through the reactor and the preheaters before it reaches its destinations in the three condensers. The columns can indirectly upset the reactor.

Path 3 is the heating and cooling circuit that starts from the reactor exit and goes through the preheaters and column reboilers. In the preheaters the hot streams give up most of their enthalpy to the incoming cold feed streams that travel back to the reactor. This path is shown

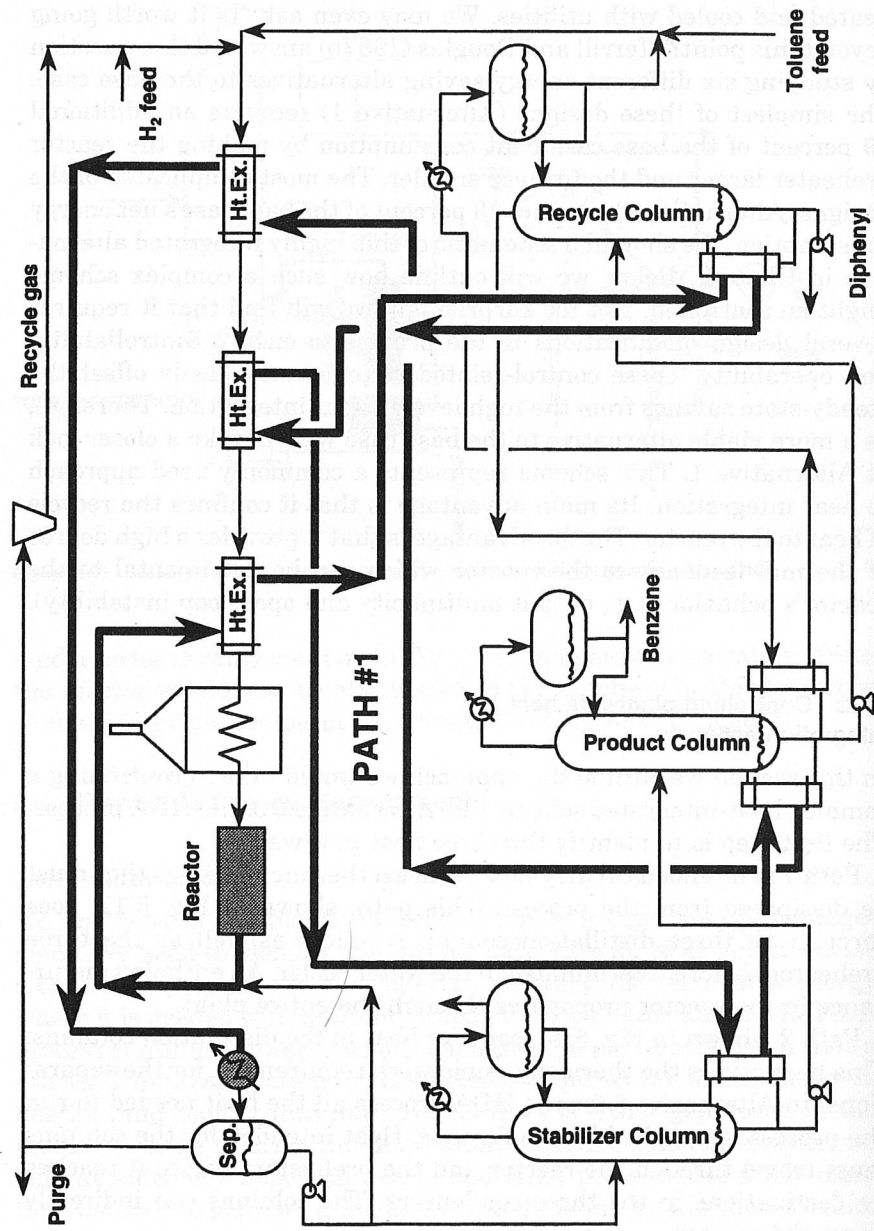


Figure 5.12 HDA process path of exothermic reaction heat.

in Fig. 5.14. Like the other two paths in this alternative, the third path also connects the reactor to the separation section thus creating the potential for interactions.

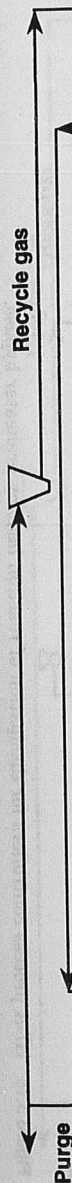
It should be clear from Figs. 5.12 to 5.14 that it would be impossible to build a process so that the flow of heat would balance itself along the three paths for all operating conditions. We could not even bank on this happening for the design conditions because of imperfections in our basic data. We need a control system to help direct the flow of heat just as we need controllers to manage material flow and inventory.

Our first and foremost concern is to dissipate the exothermic heat away from the reactor. This requires that we have the ability to bypass the feed preheaters such that excess heat from the process is not carried back with the reactor feeds. We introduce a set of bypass valves along with temperature controllers as shown in Fig. 5.15. We have elected to install the bypass valves on the cold side to lower the investment cost and facilitate maintenance. We also get better temperature control of the cold stream leaving the preheaters. In Fig. 5.15 we show the stream temperatures at the valve inlets as well as the control points. These data were obtained from the T - H diagram provided by Terrill and Douglas (1987a). Even though we bypass on the cold side, we see that one of the bypass valves has to operate at a temperature over 330°C.

Having secured the integrity of the reactor we next turn to the separation section. The concern here is to avoid picking up too much heat from the hot gas stream traveling through the reboilers. A simple solution to this problem is shown in Fig. 5.16. We have installed four more control valves in the process that bypass the reboilers. The column tray temperature controllers adjust the amount of bypass such that no more heat than necessary is delivered to each column. This design is similar to what was proposed by Terrill and Douglas (1987c).

While the control structure proposed in Fig. 5.16 satisfies the basic demands of heat management, it is not a scheme we would recommend building. First, we have reservations about the design of the recycle column reboiler and its hot bypass. As shown in Fig. 5.16, the hot gas temperature is over 425°C. When we add the fact that the operating pressure is close to 500 psia and that the stream contains mostly methane and hydrogen, we have to wonder how to design the reboiler and its bypass valve so they will operate safely and reliably.

Another concern we have is the propagation of disturbances. Consider, for example, an increase in the quenched reactor effluent temperature, T_1 (Fig. 5.17). We can use Eqs. (5.4) and (5.5) to estimate the effect that this temperature increase has on the streams around the preheater upstream of the furnace. If the bypass flow is small under normal operating conditions, we can assume that $(\dot{m}C_p)_H \approx (\dot{m}C_p)_C \approx$



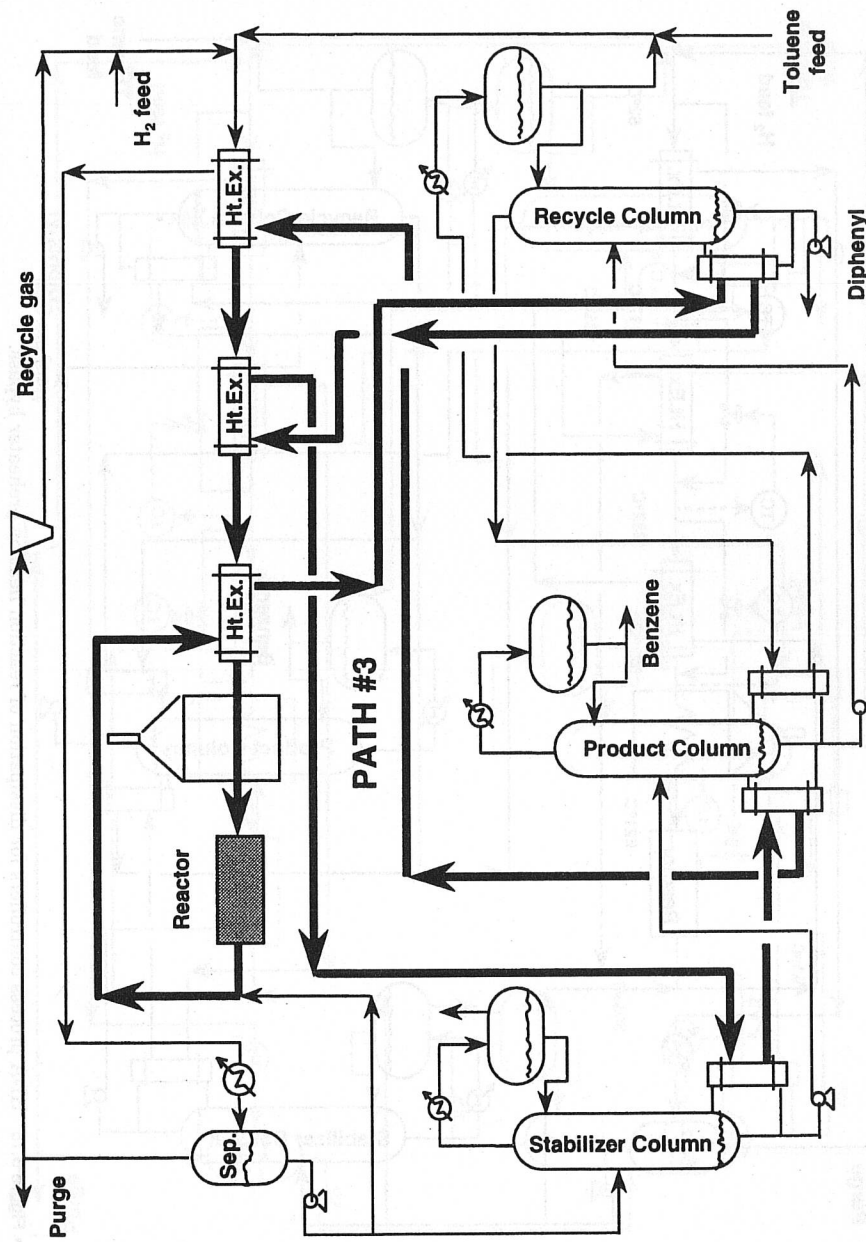


Figure 5.14 HDA process path of heat used internally for stream heating and cooling.

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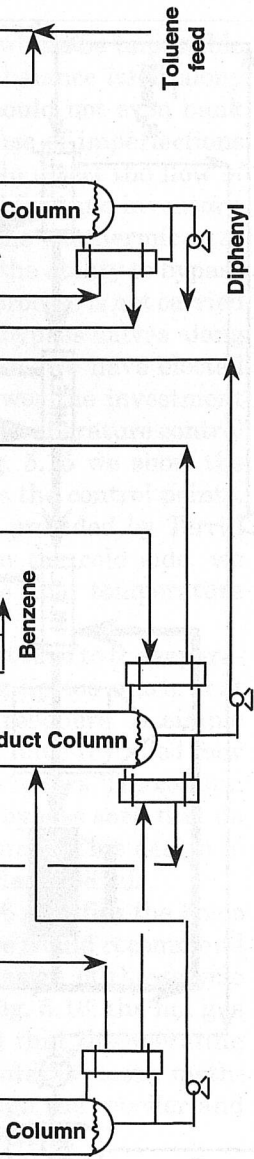


Figure 5.15 HDA process controllers for dissipation of reaction heat by preheater bypass.

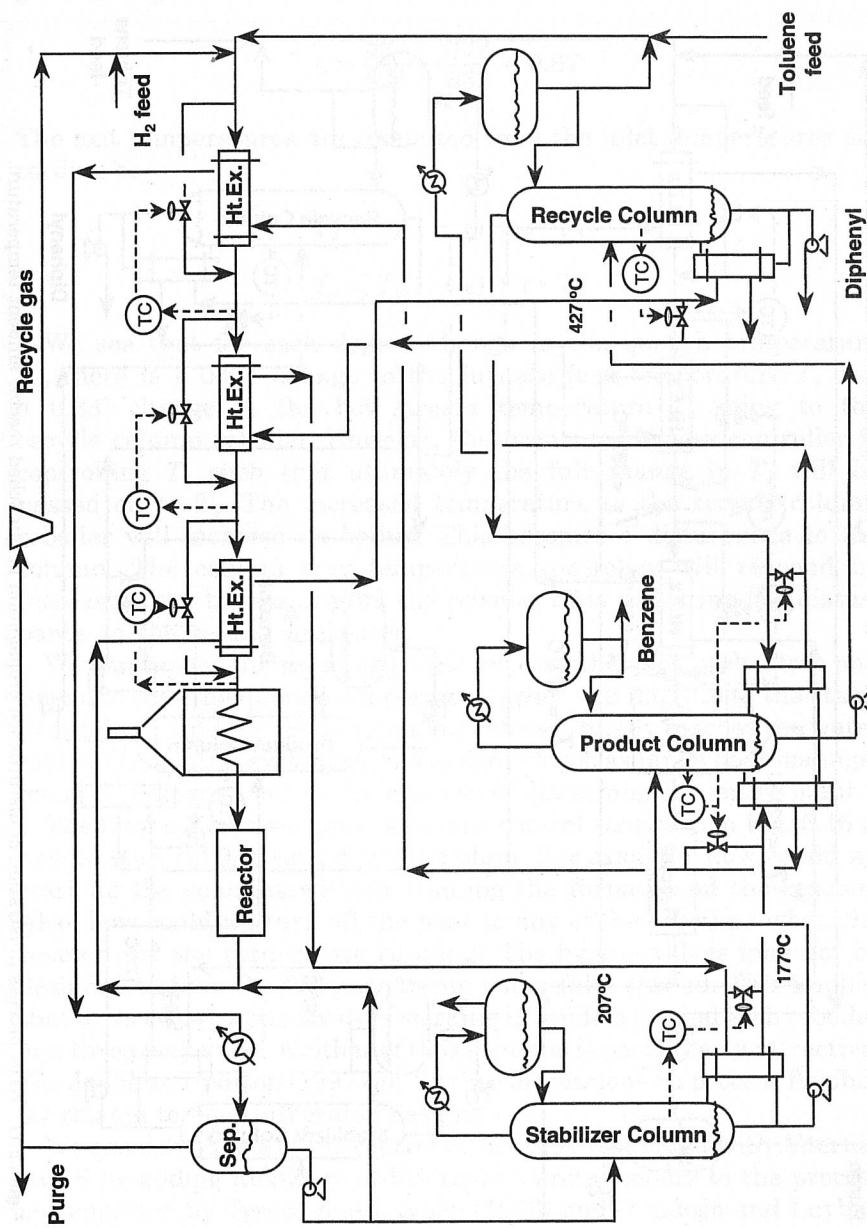


Figure 5.16 HDA process controlling heat for distillation by reboiler bypass.

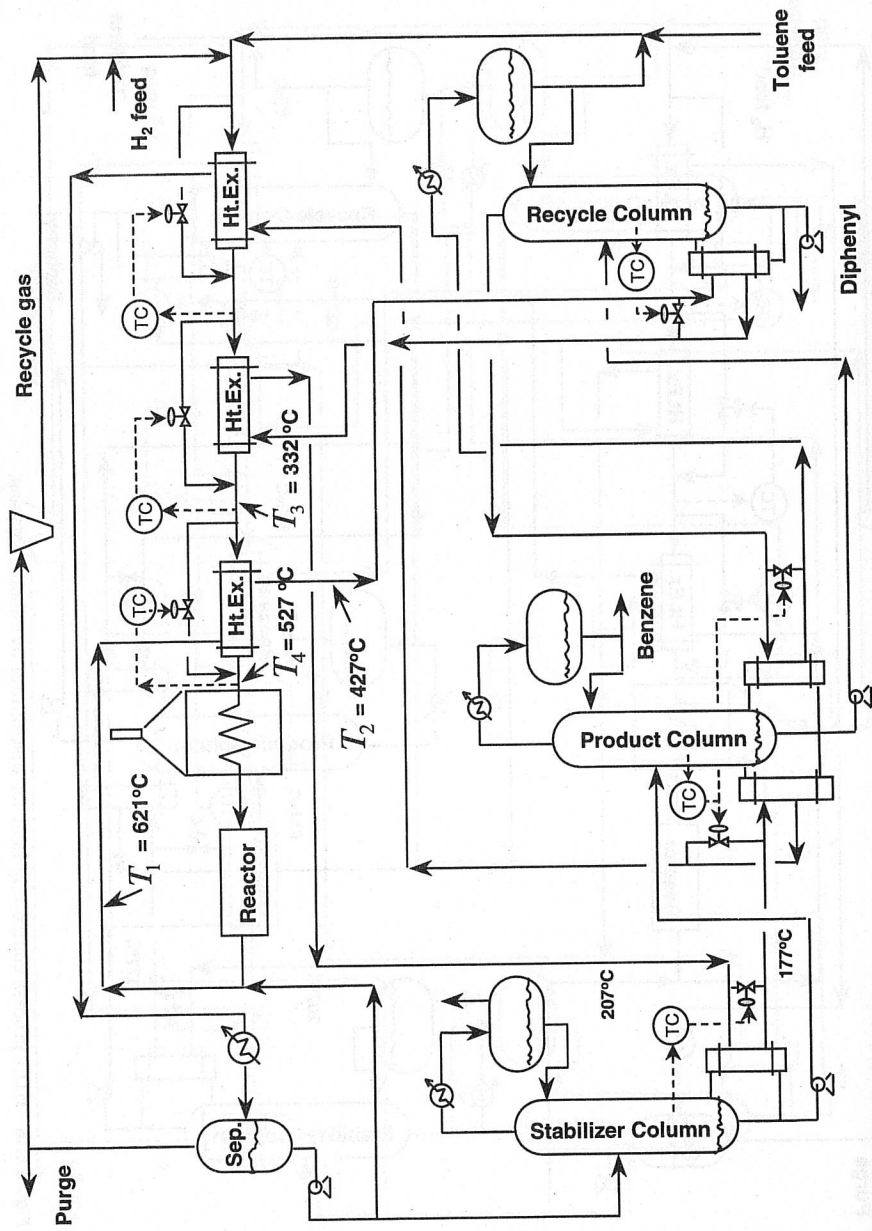


Figure 5.17 HDA process illustration of disturbance propagation in quenched reactor effluent temperature.

$(mC_p)_{\min}$. The exchanger effectiveness is then obtained from information given in Figs. 5.15 and 5.16.

$$\epsilon = \frac{527 - 332}{621 - 332} = 0.67$$

The exit temperatures are computed from the inlet temperatures according to

$$T_2 = T_1(1 - \epsilon) + \epsilon \cdot T_3$$

$$T_4 = T_3(1 - \epsilon) + \epsilon \cdot T_1$$

We see that for each degree change in the quench temperature T_1 , there is a 0.67° change in the furnace feed temperature T_4 and a 0.33° change in the hot stream temperature T_2 going to the recycle column reboiler. However, the preheater bypass controller is controlling T_4 such that ultimately the full change in T_1 will be passed on to T_2 . The increased temperature to the recycle column reboiler will increase its boilup. This becomes a disturbance to the column. The column tray temperature controller will respond by increasing the bypass around the reboiler thus deflecting the disturbance to the second preheater.

We can repeat the reasoning logic for the other two preheaters and conclude that the quench temperature upset will migrate to the stabilizer column and the product column before it finally reaches the water cooler. The control system will have done its job as far as heat management is concerned but at the expense of disturbing the entire plant.

The final concern we have about the control structure in Fig. 5.16 is how to start up and turn down the plant. For example, how would we start up the columns without running the furnace and the reactor? Also, how could we turn off the heat to any of the reboilers when the reactor and the furnace are running? The bypass valves may not be designed to take the full gas stream when fully opened. This implies that we need two control valves working in tandem around each reboiler or a three-way valve. Neither of those options is particularly attractive. See Jones and Wilson (1997) for further discussions on process flexibility related to heat integrated designs.

We can solve some of the control difficulties associated with Alternative 6 by adding auxiliary utility coolers and reboilers to the process as suggested by Tyreus and Luyben (1976) and Handogo and Luyben (1987). In Figure 5.18 we show a control configuration that uses three new reboilers and three utility coolers to improve controllability. The coolers are located in bypass streams around the process-to-process reboilers so that disturbances in the heat balance can be dissipated

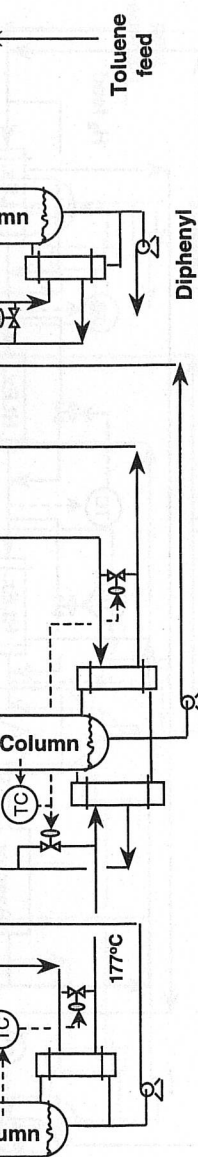


Figure 5.17 HDA process illustration of disturbance propagation in quenched reactor effluent temperature.