CHAPTER III

THEORY OF PLANTWIDE CONTROL

Plantwide contains of several connected unit operation between material recycle and energy integration make the process more performance.This chapter will offer about related to control and elements in the process. The study of MIPA process unit operation is involved in the process of many types. The devices have different control each of methods and have the effect of each unit operation. Therefore, the key principle is especially important to study carefully.

3.1 Basic Concepts of Plantwide Control

3.1.1 Buckley Basic

Buckley (1964) was studied first to suggest the idea of separating the plantwide control into two parts:

1. Material balance control
2. Production quality control

Buckley issuggested looking first at the flow of material through the system. A logical arrangement of level and pressure control loops is established, using the flow rates of liquid and gas process streams. Note that most level controllers should be proportional-only (P) to achieve flow smoothing. Then, he is proposed establishing the product-quality control loops by choosing appropriate manipulated variables. The time constants of the closed-loop product-quality loops are estimated. He tries to make these as small as possible so that good, tight control is achieved, but stability constraints impose limitations on the achievable performance.

3.1.2 Douglas doctrines

Because the cause of raw materials and the valves of products are usually much greater than the costs of capital and energy, Jim Douglas (1988) had leads to the two Douglas doctrines:

1. Minimize losses of reactants and products.
2. Maximize flow rates through gas recycle systems.

The first idea is implies that the tight control of stream compositions exiting the process to avoid losses of reactants and product. The second is rests on the principle that yield is worth more than energy. Recycles are used to improve yields in many processes. The economics of improving yields (obtaining more desired products from the same raw materials) usually overbalance the additional energy cost of driving the recycle gas compressor.

3.1.3 Drowns drill

Jim Downs (1992) are indicated the importance of looking at the chemical component balances around the entire plant and checking to see that the control structure handles these component balances effectively. All components (reactants product, and inert) have a way to leave or be consumed within the process. Most of the problems occur in the consideration of reactants, particularly when several chemical species are involved. Because raw material costs and maintain high-purity products must be minimized, most of the reactant fed into the process must be chewed up in the reactions. And the stoichiometry must be satisfied down to the last molecule. Chemical plants often act as pure integrators in terms of reactants will result in the process gradually filling up with the reactant component that is in excess. There must be a way to adjust the fresh flow rates so that exactly the right amounts of the two reactants are fed in.

3.1.4 Luyben laws

Three laws have been exploited as a result of a number of case studies of many types of systems;

1. To prevent the snowball effect, all recycle loops should be controlled flow.

2. A fresh feed stream (reactant) cannot be flow controlled unless there is essentially complete one-pass conversion of one of the reactants. This law applies to systems with reaction such as A+B →products. In systems with consecutive reactions such as A+B →M+C and M+B →D+C, the fresh feeds can be flow controlled into the system, because any imbalance in the ratios of reactants is accommodated by a shift in the amounts of the two products (M and D) that are generated. An excess of A will result in the production of more M and less D. An excess of B results in the production of more D and less M.

3. If the final product from process comes out at the top of a distillation column, the column feed should be liquid. If the final product comes out the bottom of a column, the feed to the column should be vapor. Changes in feed flow rate or feed composition have less of a dynamic effect on distillate composition than they do on bottoms composition if the feed is saturated liquid. The reverse is true if the feed is saturated vapor: bottom is less affected than distillate.

3.1.5 Richardson rule

Bob Richardson (1988) is proposed the heuristic that the largest stream should be selected to control the liquid level in a vessel. This makes good sense because it provides more muscle to achieve the desired control objective. An analogy is that it is much easier to maneuver a large barge with a tugboat than with a life. The point is that the bigger the handle you have to affect a process, the better you can control it. This is why there are often fundamental conflicts between steady-state design and dynamic controllability.

3.1.6 Shinkey schemes

Greg Shinskey (1988) is proposed a number of "advanced control" structures that permit improvements in dynamic performance. These schemes are not only effective, but they are simple to implement in basic control instrumentation. Liberal are use should be made of ratio control, cascade control, override control, and valve-position (optimizing) control.

3.1.7 Tyreus tuning

The use of P-only controllers for liquid levels, tuning of a P controller is usuallytrivial: set the controller gain equal to 1.67. This will have the valve wide open when thelevel is at 80 percent and the valve shut when the level is at 20 percent. For other controlloops, suggest the use of PI controllers. The relay-feedback test is a simple and fast wayto obtain the ultimate gain (K_u) and ultimate period (P_u). Then either the Ziegler-Nicholssettings or the Tyreus-Luyben (1992) settings can be used:The use of PID controllers should be limited to those loops where two criteria areboth satisfied:

(1) The controlled variable should have a very large signal-to-noise ratio.

(2) Tight dynamic control from a feedback control stability aspect is very crucial.

Theclassical example of the latter is temperature control in an irreversible exothermic chemical.

3.2 Integrated Processes

Three basic features of integrated chemical processes lie at the root of our need to consider the entire plant's control system: The effects of material recycle and energy integration. The explanation is chemical component inventories.

3.2.1Material recycles

Material is recycled for six basic and important reasons;

1. Increase conversion: For chemical processes involving reversible reactions, conversion of reactants to products is limited by thermodynamic equilibrium constraints. Therefore the reactor effluent by necessity contains both reactants and products. Separation and recycle of reactants are essential if the process is to be economically viable

2. Improve economics: In most systems it is simply cheaper to build a reactor with incomplete conversion and recycle reactants than it is to reach the necessary conversion level in one reactor or several in series. The simple little process illustrates

this for a binary system with one reaction A→ B. A reactor followed by a stripping column with recycle is cheaper than one large reactor or three reactors in series.

3. Improve yields: In reaction systems such as A→ B→ C, where B is the desired product, the per-pass conversion of a must be kept low to avoid producing too much of the undesirable product C. Therefore the concentration of B is kept fairly low in the reactor and a large recycle of A is required.

4. Provide thermal sink: In adiabatic reactors and in reactors where cooling is difficult and exothermic heat effects are large, it is often necessary to feed excess material to the reactor (an excess of one reactant or a product) so that the reactor temperature increase will not be too large. High temperature can potentially create several unpleasant events: it can lead to thermal runaways, it can deactivate catalysts, it can cause undesirable side reactions, it can cause mechanical failure of equipment, etc.

5. Prevent side reactions: A large excess of one the reactant is often used so that the concentration of the other reactant is kept low. If this limiting reactant is not kept in low concentration, it could react to produce undesirable products. Therefore the reactant that is in excess must be separated from the product components in the reactor effluent stream and recycled back to the reactor.

6. Control properties: In many polymerization reactors, conversion of monomer is limited to achieve the desired polymer properties. These include average molecular weight, molecular weight distribution, degree of branching, particle size, etc. Another reason for limiting conversion to polymer is to control the increase in viscosity that is typical of polymer solutions. This facilitates reactor agitation and heat removal and allows the material to be further processed.

3.2.2 Energy integration

The fundamental reason for the use of energy integration is to improve the thermodynamic efficiency of the process. This translates into a reduction in utility cost. For energy-intensive processes, the savings can be quite significant.

3.2.3 Chemical component inventories

We can characterize a plant's chemical species into three types: reactants, products and by-product. A material balance for each of these components must be satisfied. This is typically not a problem for products and inert. However, the real problem usually arises when we consider reactants (because of recycle) and account for their inventories within the entire process. Every molecule of reactants fed into the plant must either be consumed via reaction or leave as an impurity or purge.

Because of their value, we want to minimize the loss of reactants exiting the process since this represents a yield penalty. So we prevent reactants form leaving. This means we must ensure that every mole of reactant fed to the process is consumed by the reactions.This is an important concept and is generic to many chemical processes. From the viewpoint of individual units, chemical component balancing is not a problem because exit streams form the units automatically adjust their flows and compositions. However, when we connect units together with recycle streams, the entire system behaves almost like a pure integrator in terms of the reactants. If additional reactant is fed into the system without changing reactor conditions to consume the reactant, this component will build up gradually within the plant because it has no place to leave the system.

Plants are not necessarily self-regulating in terms of reactants. We might expect that the reaction rate will increase as reactant composition increases.However, in system with several reactants (e.g., A + B →products), increasing one reactant composition will decrease the other reactant composition with an uncertain net effect on reaction rate.

3.3 The Plantwide Control Problem

3.3.1 Units in Series Problem

If process units are arranged in a purely series configuration, where the products of each unit feed downstream units and there is no recycle of material or energy, the plantwide control problem is greatly simplified. We do not have to worry about the issues discussed in the previous section and we can simply configure the control scheme on each individual unit operation to handle load disturbances.

If production rate is set at the front end of the process, each unit will only see load disturbances coming from its upstream neighbor. If the plant is set up for "on-demand" production, changes in throughput will propagate back through the process. So any individual unit will see load disturbances coming from both its downstream neighbor (flow rate changes to achieve different throughputs) and its upstream neighbor (composition changes as the upstream units adjust to the load changes they see).

Figure 3.1: compares these two possible configurations for a simple plant. A fresh stream containing a mixture of chemical components A, B and C is feed into a two-column distillation train. The relative volatilities are $\alpha_A > \alpha_B > \alpha_C$, and we select the "direct" (or "light-out-first) separation sequence: A is taken out the top of the first column and B out the second column.

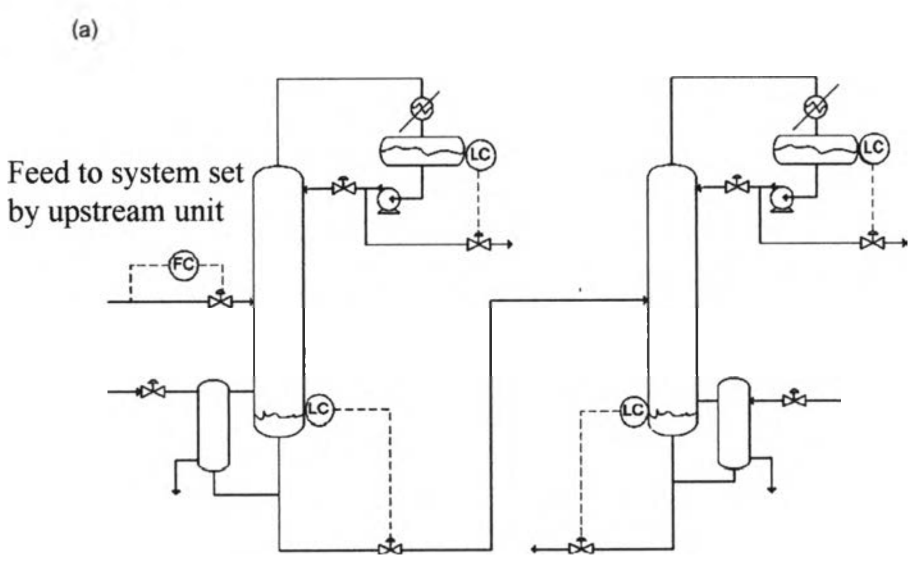


Figure 3.1: Units in series: Level control in direction of flow

3.3.2 Effects of Recycle

Most real process contains recycle stream. In this case,the problem is not an issue that is clearly expressed and not expressed by supposedly took place. Two basic effect of recycle is: Recycle has an impact on the dynamic of the process. The overall time constant can be much different than the sum of the time constants of the time constants of the individual units. Recycle stream leads to the "snowball effect". A small change in throughput of feed composition can lead to a large change in steady-state recycle streams flow rates.

3.3.3 Snowball effect

Snowball effect is high sensitivity of the recycle flow rates to small disturbances. When feed conditions are not very different, recycle flow rates increase drastically, usually over a considerable period of time. Often the equipment cannot handle such a large load. It is a steady-state phenomenon but it does have dynamic implications for disturbance propagation and for inventory control.

The large swings in recycle flow rates are undesirable in plant because they can overload the capacity of separation section or move the separation section into a flow region below its minimum turndown. Therefore it is important to select a plantwide control structure that avoids this effect.

3.3.4. Reaction and Separation Section Interaction

For the process considered in the previous section where the reaction is $A \rightarrow B$, the overall reaction rate depends upon reactor holdup, temperature (rate constants), and reactant composition (mole fraction A) $R=V_R\ k_Z$. The two control structures considered above produce fundamentally different behavior in handling disturbances. In the first, the separation section must absorb almost all of the changes. This translates into a very significant change in the composition of the feed stream to the separation section. This means the load on the separation section changes significantly, producing large variations in recycle flow rates.

In the second structure, both reactor holdup and reactor composition z can change, so the separation section sees a smaller load disturbance. This reduces the magnitude of the resulting change in recycle flow because. The effects of the disturbance can be distributed between the reaction and separation sections.

A very useful general conclusion from this simple binary system can be depicted that is applicable to more complex processes: changes in production rate can be achieved only by changing conditions in the reactor. This means something that affects reaction rate in the reactor must vary: holdup in liquid-phase reactor, pressure in gas-phase reactors, temperature, concentrations of reactants, and catalyst activity or initiator addition rate. Some of these variables affect the conditions in the reactor more than others. Variables with a large effect are called dominant. By controlling the dominant variables in a process, partial control is achieved.

The term partial control arises because it typically haves fewer available manipulators than variables that would like to control. The set points of the partial control loops are than manipulated to hold the important economic objectives in the desired ranges.

Hence a goal of the plantwide control strategy is to handle variability in production rate and in fresh reactant feed compositions while minimizing changes in the feed stream to the separation section. This may not be physically possible or economically feasible. But if it is, the separation section will perform better to accommodate these changes and to maintain product quality, which is one of the vital objectives for plant operation. Reactor temperature, pressure, catalyst/initiator activity, and holdup are preferred dominant variables to control compared to direct or indirect manipulation of the recycle flows, which of course affect the separation section.

3.3.5 Single Composition Control on Distillation Control

Here the composition of one product is controlled while the composition of the other product is allowed to float. In the chemical industry over 90% of the columns are operated under single composition control compared to dual composition control, which controls both the overhead and the bottoms product compositions.

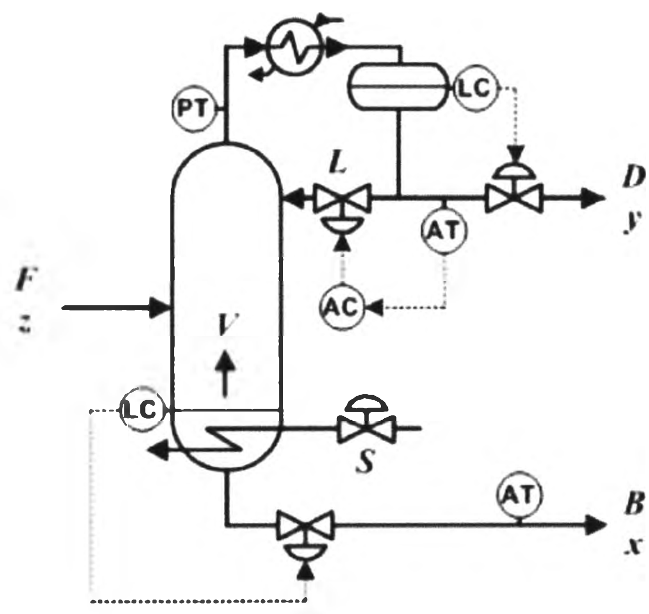


Figure 3.2: Single composition control using L to control the purity of the overhead product. The symbol for a control valve represents a flow control loop.

Figure 3.2: shows single composition control are using the reflux to control the purity of the overhead product while maintaining a fixed reboiler duty. The bottom composition is not controlled directly and can vary significantly as the feed composition changes. The control performance of the overhead product is generally best when reflux, L, rather than either the distillate flow rate, D, or the reflux ratio, L/D is the MV. L is the fastest-acting MV for the overhead and the least sensitive to feed composition changes. Because the reboiler duty is fixed, coupling is not an issue

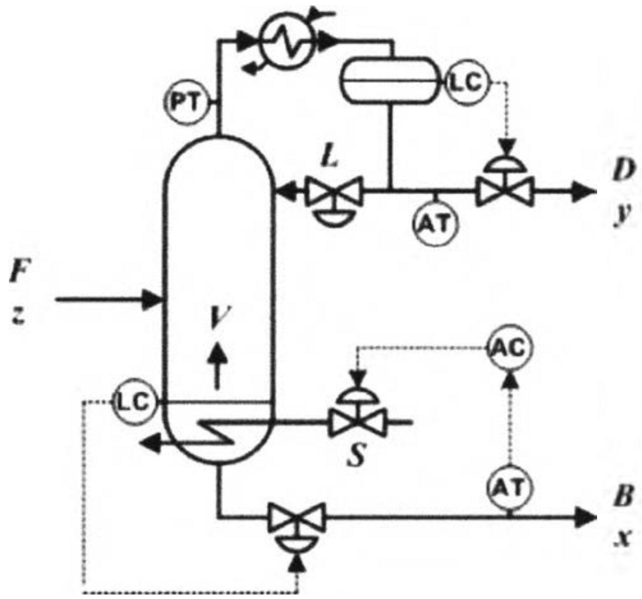


Figure 3.3: Single composition control using V to control the purity of the bottom product .The symbol for a control valve represents a flow control loop.

When the bottom product is controlled by single composition control, the control configuration shown in Figure 3.3 is recommended. Because the boilup rate, V, is faster acting and less sensitive to disturbances than either the bottoms product rate, B, or the boilup ratio, V/B. V is used to control the bottom product composition with the reflux rate fixed, which allows the overhead composition to float. Single composition control is much easier to implement, tune and maintain than dual composition control. The choice between single and dual composition control is based on the tradeoff between the additional cost associated with dual composition control (analyzer costs, increased controller maintenance, etc.) and the economic benefit of dual composition control (increased product recovery and reduced utility costs). While single composition control is in widespread use in the chemical industry, dual composition control is generally preferred for refinery columns and columns that produce high-volume chemical intermediates because the energy usage for these columns is much larger.

3.4 Step of Plantwide Process Control Design Procedure.

The nine steps of the design procedure center around the fundamental principles of plantwide control: energy management, production rate, product quality, operational, environmental, and safety.

Step 1: Establish control objectives.

Assess the steady-state design and dynamic control objectives for the process.

This is probably the most important aspect of the problem because different control objectives lead to different control structures. The "best" control structure for a plant depends upon the design and control criteria established.

These objectives include reactor and separator and separation yields, product quality specifications, product grades and demand determination, environmental restrictions, and the range of safe operation conditions.

Step 2: Determine control degrees of freedom

Count the number of control valves available.

This is the number of degrees of freedom for control, i.e., the number of variables that can be controlled to set point. The valves must be legitimate (flow through a liquid-filled line can be regulated by only one control valve). The placement of these control valves can sometimes be made to improve dynamic performance, but often there is no choice in their location.

Most of these valves will be used to achieve basic regulatory control of the process;

(1) Set production rate
(2) Maintain gas and liquid inventories
(3) Control product qualities
(4) Avoid safety and environmental constraints.

Any valves that remain after these vital tasks have been accomplished can be utilized to enhance steady-state economic objectives or dynamic controllability (e.g., minimizes energy consumption, maximize yield, or reject disturbances).

Step3: Establish energy management system

Make sure that energy disturbances do not propagate throughout the process by transferring the variability to the plant utility system.

The term energy management is used to describe two functions:

(1) To provide a control system that removes exothermic heats of reaction form the process. If heat is not removed to utilities directly at the reactor, then it can be used elsewhere in the process by other unit operations. This heat however must ultimately be dissipated to utilities.

(2) To provide a control system that prevents the propagation of thermal disturbances and ensures the exothermic reactor heat is dissipated and not recycled.

Process-to-process heat exchangers and heat-integrated unit operations must be analyzed to determine that there are sufficient degrees of freedom for control.

Heat removal in exothermic reactors is crucial because of the potential for thermal runaways. In endothermic reactions, failure to add enough heat simply results in the reaction slowing up. If the exothermic reactor is running adiabatically, the control system must prevent excessive temperature rise through the reactor (e.g., by setting the ratio of the flow rate of the limiting fresh reactant to the flow rate of a recycle stream acting as a thermal sink).

Heat transfer between process streams can create significant interaction. In the case of reactor feed/effluent heat exchangers it can lead to positive feedback and even instability. Where there is partial condensation or partial vaporization in a process-to-process heat exchanger, disturbances can be amplified because of heat of vaporization and temperature effects.

Step 4: set production rate

Establish the variables that dominate the productivity of the reactor and determine the most appropriate manipulator to control production rate.

Throughput changes can be achieved only by altering, either directly or indirectly, conditions in the reactor. To obtain higher production rates, the overall reaction rates must be increased. This can be accomplished by raising temperature (higher specific reaction rate), increasing reactant concentrations, increasing reactor holdup (in liquid-phase reactors), or increasing reactor pressure (in gas-phase reactors).

Our first choice for setting production rate should be to alter one of these variables in the reactor. The variable that is selected must be dominant for the reactor. Dominant reactor variables always have significant effects on reactor performance. For example, temperature is often a dominant reactor variable. In irreversible reactions, specific rates increase exponentially with temperature. As long as reaction rates are not limited by low reactant concentrations, temperature can be increased to increase production rate in the plant. In reversible exothermic reactions, where the equilibrium constant decreases with increasing temperature, reactor temperature may still be a dominant variable. If the reactor is large enough to reach chemical equilibrium at the exit, the reactor temperature can be decreased to increase production.

There are situations where reactor temperature is not a dominant variable or cannot be changed for safety or yield reasons. In these cases, another dominant variable must be found, such as the concentration of the limiting reactant, flowrate of initiator or catalyst to the reactor, reactor residence time, reactor pressure, or agitation rate.

Once the dominant variables must be identified, the manipulators (control valves) must also be identified that are most suitable to control them. The manipulators are used in feedback control loops to hold the dominant variables at set point. The set points are then adjusted to achieve the desired production rate, in addition to satisfying other economic control objectives.

Whatever variable is chosen, it can provide smooth and stable production rate transitions and to reject disturbances. A variable that has the least effect on the separation section but also has a rapid and direct effect on reaction rate in the reactor without hitting an operational constraint is often wanted select.

When the set point of a dominant variable is used to establish plant production rate, the control strategy must ensure that the tight amounts of fresh reactants are brought into the process. This is often accomplished through fresh reactant makeup control based upon liquid levels or gas pressures that reflect component inventories.

However, design constraints may limit our ability to exercise this strategy concerning fresh reactant makeup. An upstream process may establish the reactant feed flow sent to the plant. A downstream process may require on-demand production, which fixes the product flowrate from the plant. In these cases, the development of the control strategy becomes more complex because the set point of the dominant variable on the basis of the production rate that has been specified externally must be somehow adjusted. The production rate with what has been specified externally must be balanced. This cannot be done in an open-loop sense. Feedback of information about actual internal plant conditions is required to determine the accumulation or depletion of the reactant components.

Step 5: Control product quality and handle safety, operational, and environmental constraints.

Select the “best” valves to control each of the product-quality, safety, and environmental variables. The tight control of these important quantities for economic and operational reasons is wanted. Hence the manipulated variables such that dynamic relationships between the controlled variables and manipulated variables, feature small time constants and dead times and large steady-state gains should be selected. The former gives small closed-loop time constants and the latter prevents problems with the range ability of the manipulated variable (control valve saturation).

It should be noted that establishing the product-quality loops first. Before the material balance control structure, is a fundamental difference between our plantwide control design procedure and Buckley's procedure, since product quality considerations have become more important is recent year, this shift in emphasis follows naturally.

The magnitudes of various flow rates also come into consideration. For example, temperature (or bottoms product purity) in a distillation column is typically controlled by manipulating stem flow to the reboiler (column boil up) and base level is controlled with bottoms product flow rate. However, in columns with a large boil up ratio and small bottoms flow rate, these loops should be reversed because boil up has a larger effect on base level than bottoms flow (Richardson rule). However, inverse response problems in some columns may occur when base level is controlled by heat input. High reflux ratios at the top of a column require similar analysis in selecting reflux or distillate to control overhead product purity.

Step 6: Control Inventories (Pressures and Levels) and Fix a Flow in Every Recycle Loop.

Determine the valve to control each inventory variable. These variables include all liquid levels and gas pressures. An inventory variable should typically be controlled with the manipulated variable that has the largest effect on it within that unit.

Proportional-only control should be used in nonreactive level loops for cascaded units in series. Even in reactor-level control, proportional control should be considered to help filter flow-rate disturbances to the downstream separation system. There is nothing necessarily sacred about holding reactor level constant.

In most processes a flow controller should be present in all liquid recycle loops. This is a simple and effective way to prevent potentially large changes in recycle flow that can occur if all flows in the recycle loop are controlled by levels. Two benefits result from this flow-control strategy. First, the plant's separation section is not subjected to large load disturbance. Second, consideration must be given to alternative fresh reactant makeup control strategies rather than flow control.In a dynamic sense, level controlling all flows in a recycle loop is a case of recycling disturbances and should be avoided. Gas recycle loops are normally set at maximum circulation rate, as limited by compressor capacity, to achieve maximum yields.

Step 7: Check Component Balances.

Identify how chemical components enter, leave, and are generated or consumed in the process.

Ensure that the overall component balances for each chemical species can be satisfied either through reaction or exit streams by accounting for the component's composition or inventory at some point in the process. Light, intermediate, and heavy inert components must have an exit path from the system. Reactant must be consumed in the reaction section or leave as impurities in the product streams. Fresh reactant makeup feed streams can be manipulated to control reactor feed composition or a recycle stream composition (or to hold pressure or level as noted in the previous step). Purge streams can also be used to control the amount of high- or low-boiling impurities in a recycle stream.

Component balances can often be quite subtle. They depend upon the specific kinetics and reaction paths in the system. They often affect what variable can be used to set production rate or rate in the reactor.

Step 8: Control Individual Unit Operations

Establish the control loops necessary to operate each of the individual unit operations.

Many effective control schemes have been established over the years for chemical units. For example, a tubular reactor usually requires control of inlet temperature. High-temperature endothermic reactions typically have a control system to adjust fuel flow rate to a furnace supplying energy to the reactor. Crystallizers require manipulation in the stack gas from a furnace is controlled to prevent excess fuel usage. Liquid solvent feed flow to an absorber is controlled as some ratio.

Step 9: Optimize Economics or Improve Dynamic Controllability

Establish the best way to use remaining control degrees of freedom.

The after is satisfying all of the basic regulatory requirements. Additional degrees of freedom involving control valves that have not been used and set points in some controllers that can be adjusted. These can be used either to optimize steady-state economic process performance or to improve dynamic response.

3.5 Plantwide Energy Management

3.5.1. Heat Exchanger Dynamics

Heat Exchangers have fat dynamics compared to other unit operations in a process. Normally the time constant is measured in second but could be up to a few minutes for large exchangers. Process-to-process exchangers should be modeled rigorously by differential equations since they are distributed systems. This introduces the correct amount of dead time and constants in exit stream temperatures, but the models are inconvenient to solve.

For propose 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 heat exchanger.

3.5.2. Heat Pathways

The most of energy are required for heating certain streams within the process is matched by similar amount required for cooling other streams. Heat recover from cooling streams 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).

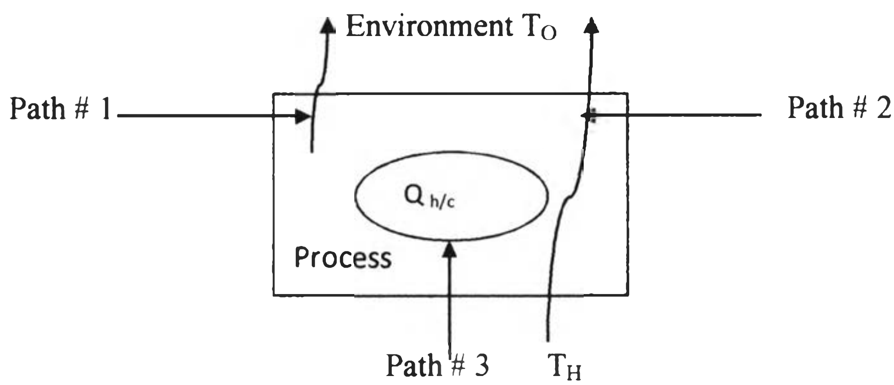


Figure 3.4: Heat Pathways

From a plantwide perspective we can now discern 3 different "heat pathway" in the process. See Figure 3.4 the first pathways dissipate to the environment heat generated by exothermic reaction and by degradation of mechanical work (e.g. compression, pressure drop, and friction).This pathway is from inside the process and flow out. It is of course possible to convert some of the heat to work as it is from high temperature in the process.

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

The third pathway is internal to 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.

3.5.3. Heat Recovery

We can make great improvements in the plant's thermal efficiency by recycling much of the energy needed for heating and cooling process streams.

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 dynamic and control of a plant and how we can manage energy in plants with a high degree of heat recovery.

3.5.4 Heat Exchanger Network

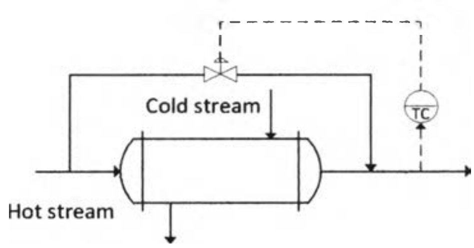
It is generally accepted that an optimal network must feature a minimum number of units that reflects on a capital cost and minimum utility consumption that reflects on operating costs. A good engineering design must exhibit minimum capital and operating costs. For Heat Exchanger Network (HEN) synthesis, safety, etc. in recent years the attention in HEN synthesis has been focused on the operability features of a HEN, e.g. the ability of a HEN to tolerate unwanted changes in operating condition. It has been learned that considering only a cost objective in synthesis may lead to a worse network, i.e. a minimum cost network may not be operable at some neighboring operating condition. The design must not only feature minimum cost, but also be able cope with a fluctuation or changes in operating condition. The ability of a HEN to tolerate unwanted changes is called resiliency. It should be note that the ability of a HEN to tolerate wanted changer is called flexibility.

The resiliency property of a design becomes an important feature to be accounted for when the extent of integration of a design introduces significant interactions among process components. The energy integration of a HEN generates a quite complex interaction of process stream, despite the fact that transfer of heat from hot to clod process streams is the only activity of the network. The goal of a network is to deliver the process streams to their target temperatures by using most of their heating and cooling availability and a minimum of heating and cooling utilities. The process streams are coupled through a net of heat exchangers. Changes in conditions of one stream in the network may affect the performances of many heat exchanges and the conditions of several process teams. Since resiliency is a property of a network structure.

3.5.5 Control of Utility Exchangers

The purpose of unit operation controls is to regulate the amount of energy supplies or removed. This is typically done by measuring a temperature in the process and manipulating the flow rate of utility. A PI-controller is adequate in most instances. The location of the temperature measurement depends upon the purpose of the heat exchange. The control point should be location where then effects of the added energy are felt the most. When the utility exchanger is used for stream heating and cooling, the control point is on the stream being heated or cooled.

Use of Auxiliary Exchanger: When the bypass method is used for unit operation control, we have several choices about the bypass location and the control point. Figure 3.5 shows the most common alternatives. For choosing the best option, it depends on how we define the beat. Design consideration 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 consideration would also suggest a small bypass flow to minimize the exchanger and control valve sizes.



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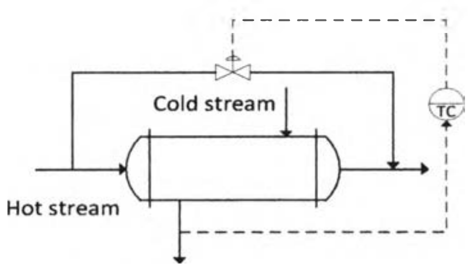

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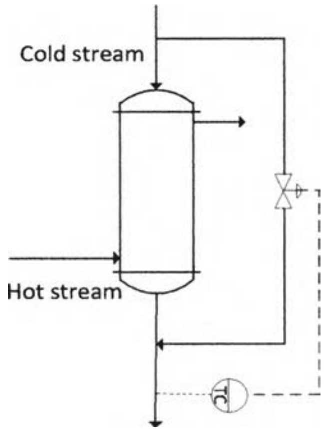

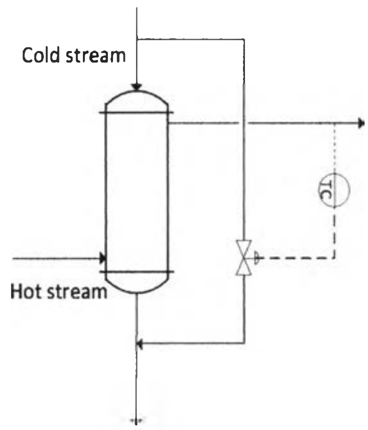

C D

Figure 3.5: Bypass control of process to process heat exchangers where:

(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.

From a control standpoint, we should measure the most important stream, regardless of temperature, and bypass on the same side as well we control (see Figure 3.5 A 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 streams. 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.

3.6 Plantwide Process Control

Control analysis and control system are designed for chemical and petroleum processes have traditional followed the "unit operations approach". First, all of the control loops were established individually for each unit or piece of equipment in the plant. Then the pieces were combined together into an entire plant. This meant that any conflicts among the control loops somehow had to be reconciled. The implicit assumption of this approach was that the sum of the individual parts could effectively comprise the whole of the plant's control system.

Over the last few decades, process control researchers and practitioners have developed effective control schemes for many of the traditional chemical unit operations. And for processes where these unit operations are arranged in series, each downstream unit simply sees disturbances from its upstream neighbor.

Most industrial processescontain a complex flow sheets with several recycle streams, energy integration, and many different unit operation. Essentially, the plantwide control problem is how to development the control loops needed to operate an entire process and achieve its design objective. Recycle stream and energy integration introduce a feedback of material and energy among unit upstream and downstream. They also interconnect separate unit operations and create a pate for disturbance propagation. The resent of recycle streams profoundly alters that is not localized to an isolate part of the process.

Despite this process complexity, the unit operation to control system design has worked reasonably well. In the past, plants with recycle streams contained many surge tanks to buffer disturbance, to minimize interaction, and to isolate unit in the sequence of the material flow. This allowed each unit to be controlled individually. Prior to the 1970s, low energy cost meant little economic incentive for energy integration. However, there growing pressure to reduce capital environmental concerns. This has prompted design engineers to start eliminating many surge tanks, increasing recycle stream, and introducing heat integration for both exiting and plant. Often this is done without a complete understanding of their effects on plant operability.

So economic forces within the chemical industry are compelling improved capital productivity, requirements for on-aim product quality control grow increasingly tighter. More energy integration occurs. Improved yields, which reduce raw material costs, are achieved via lower reactant per-pass conversion and higher material recycles rates through the process. Better product quality, energy integration, and higher yields are all economically attractive in the steady-state flow sheets by the present significant challenges to smooth dynamic plant operation. Hence an effective control system regulating the entire plants operation and process designed with good dynamic performance play critical parts in achieving the business objectives of reducing operating and capital costs.

Buckley (1964) proposed a control design procedure for the plantwide control problem that consisted of two stages. The first stage determined the material balance control structure to handle vessel inventories for low-frequency disturbances. The second established the product quality control structure to regulate high-frequency disturbances. This procedure has been widely and effectively utilized. It has served as the conceptual framework in many subsequent ideas for developing control systems for complete plants. However, the two-stage Buckley procedure provides little guidance concerning three important aspects of plantwide control strategy. First, it does not explicitly discuss energy management. Second, it does not address the specific issues of recycle systems. Third, it does not deal with component balance in the context of inventory control. By placing the priority on material balance over product quality control, the procedure can significantly limit the flexibility in choosing the latter.

The goals for an effective plantwide process control system include.

1. Safe and smooth process operation.
2. Tight control of product quality in the face of disturbances.
3. Avoidance of unsafe process condition.
4. A control system runs in automatic, not manual, requiring minimal operator attention.
5. Rapid rate and product quality transitions.
6. Zero unexpected environmental releases.

3.6.1 New Plantwide Control Structure Design Procedure (Wongsri, 2012)

The procedure consists of eight steps of the plantwide control structure design which makes use of chemical engineering and process knowledge and heuristics. The design of plantwide control structure should be viewed *as a whole; taking into consideration of the whole plant.* The design step called plantwide level design is the decision of how to regulate the whole plant albeit a single entity as smoothly as possible. Then the designs of control loops that locally function are handled at the unit level design.

In the plantwide level, we establish a fixture plant: creating a material-balanced process plant. In establishing a fixture plant we do these: keep the raw materials entered to the reactor fixed, adjust the flow of exit material streams (products, by-products, and inert) according to their accumulations, and locate the quantifiers for the rest of the components to design the control loops to regulate their inventories.

For operating a plant as smoothly as possible wemust diminish the disturbance by designing the control loops to reject or direct the disturbances taking a whole plant into consideration. We also use the heat disturbances management principle i.e. we will reject any not directly related to quality to the environment via the next and nearest exit points. We make use of the material path way analysis to design the control loops to direct the material disturbances.

At the unit level, we design the control loops of the rest of units that involve pressures, flow rates or levels, for examples, pumps and compressors.

We are evaluate the performance of the control structure designed using the industrial dynamic simulator.

If the opportunities for heat integration and process modification exist, we later are designed the control structures of the heat integrated plant (HIP) and modified plant. The control structure of the new designs then will be compared with the previous designs. The trade-off between cost savings and control performances will be discussed.

In this research, the plantwide control structures of the ammonia production are designed using Wongsri procedure. The design procedure is carried out in eight steps as follow:

Step 1: Gather relevant plant information and control objective including constraints for control.Before initiating work on the control structure design, is it necessary to obtain all information relevant to process control. The process objectives and constraints will determine the lower/upper bounds on the control variables as well as set points on quality variables.

Step 2: List manipulated variables (control degree of freedom, CDOF).

The CDOF can be obtained using the guideline given in table 2 and the guideline for pairing the controlled variables with the manipulated variables.

Table 3.1 Degree of freedom for simple units

Unit	DOF / unit
Independent stream	1
Heater, cooler, pump, and compressor	1
Heat exchanger with a by-pass stream	1
Adiabatic plug flow reactor	0
Non-adiabatic plug flow reactor	1
Adiabatic flash separator	2
Simple distillation column	5

Step 3: Establish fixture plant.

The principal idea of establishing a fixture plant is first to have an entire plant fluid-filled and a material-balanced. This idea is similar to creating *hydraulic* control structure proposed by Buckley [7]. It does not by all mean that all material balances control is managed at this step. Only the control loops conforming component balances are designed. In this step, we provide handles to regulate material flows and therefore eliminating the snowball effect or material accumulation. By establishing a fixture plant we mean creating a material-balanced process plant:

(3.1) Keep the materials entered and reentered fixed.

$$q_i(t)+q_r(t)=\text{constant} \tag{1}$$

This leaves the recycle streams free to adjust; one degree of freedom is added to the process.

If the composition of the recycle streams differ from the fresh feed stream significantly, each stream are separately controlled:

$$q_i(t) = \text{constant} \tag{2}$$

$$q_r(t) = \text{constant} \tag{3}$$

In this settlement, the flow of recycle stream cannot be used to regulate, e.g., the level of the reflux drum.

(3.2) Adjust the flow of exit material streams (products, by-products, and inert) according to their accumulations.

$$q_o(t) = q_i(t)\text{-}dq/dt \tag{4}$$

(3.3) Locate the quantifiers; i.e. the indicators of the representative accumulation, for the rest of the components and design the control loops to regulate their inventories in the plant. The quantifier can be volume (mass), pressure, or flow rate.

$$q_p(t) = \text{-}dq/dt \tag{5}$$

$$q_p(t) = \text{constant} \tag{6}$$

In retrospect, the material balances are checked in this step, since the control loops generated accomplish the plantwide material balances. Therefore, it is guaranteed the plantwide inventory will be regulated.

Step 4: Handling the disturbances.

In this step, the disturbances are handled by configuring the control loops employing the principle of disturbances management:

(4.1) Heat Disturbances

The Heat disturbance is divided into 2 categories. Heat Disturbance Category 1 (HDC1) is the heat disturbance that does not directly effect on product qualities. The heat disturbance in a process stream is usually rejected at a point before a reactor or separator by a heater or a cooler to maintain the stream temperature at its desired value. Heat Disturbance Category 2 (HDC2) is the heat disturbance that will affect the product qualities, such as heat disturbance in a process stream toward to a reactor or a separator. The disturbance is manipulated to regulate the composition.

(4.1.1) Direct the thermal disturbances that are not directly related to quality to the environment via the next and nearest exit points, usually heaters or coolers, to keep the thermal conditions of process stream fixed. The thermal condition of process stream is changed along the process plant, usually by heater or cooler of process to process heat exchanger.

(4.1.2) Manage the thermal disturbance that related to quality in order to maintain the product specification constraints.

(4.2) Material disturbances

The configuration of the control loops depend on the desired material pathways. The pathways can be obtained by analyzing the results of the material disturbance tests. The test is suggested to be done on the changing of composition, total flow, and component flow. The material disturbances can be generated at reactors and separators, besides coming with feeds and recycle streams. So if the feeds and recycle streams are fixed, the only places that alter the material (total or component) flow rates are the reactors and the separators. At reactor, its inlet temperature is adjusted in order to keep the reactor component flow rate or its composition in outlet stream. The decision of whether how to choose to control the component flow or the composition or not to control is based on the profit maximization or the smooth operation policies.

Since the distillation columns, usually the one-point control is common. To control top or bottom temperatures, depend on the material disturbance rejection policy.

Step 5: Design the control loops for the remaining control variables and/or adding enhanced controls, i.e. cascade, feed forward controls.

Step 6: Energy management via heat exchanger networks.

If potential heat exchanger networks or alternative heat integrated processes (HIPs) exist, list additional control variables and manipulated variables.

Step 7: Optimize economics and/or improve control performance. For example, the controls scheme/structure of the reactor (e.g. temperature/composition sensor location), the control scheme of the distillation column (e.g. reflux to feed ratio control), the optimal operating temperatures of the reactors, the recycle flow rates, the sequence of separation, etc. If the opportunity of optimization exists, we might backtrack to the previous step as dictated.

Step 8: Validate the designed control structures by rigorous dynamic simulation.

The measures can be costs, raw material and energy consumptions, control performances of the total plant or some selected loops, etc.