CHAPTER III

Theory of Plantwide Control

3.1 The main functions of control system

In general, the control system installed in process has three main function.

3.1.1 To reject disturbance It is the main objective in installing control system. The external disturbance is uncertain so the operator cannot monitor the changing in process. As a result, the control system must be installed to follow the changing of the process and manipulate the process variable to compensate for the disturbance from external factors.

3.1.2. To maintain stability The stability is necessary for every process. As a result the control system is set to improve the process stability for the guarantee of quality of product, safety to equipment of process and plant.

3.1.3. To keep the process performing highest efficiency Besides rejecting disturbance and maintaining stability, the control system can achieve the great profit because it losses less energy and raw materials during the operating. Moreover the product will meet the required specification and have high production rate.

3.2 Integrated Processes

Three basic features of integrated chemical processes lie at the root of the need to consider the entire plant's control system:

(1) The effect of material recycle

(2) The effect of energy integration

(3) The need to account for chemical component inventories.

If these issues were not had to worry about, then a complex plantwide control problem was not had to deal with. However, there are fundamental reasons why each these exists in virtually all real processes.

3.2.1. Material recycle

Material is recycle for six basic and important reasons.

- 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.

- 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. A reactor followed by a stripping column with recycle is cheaper than one large reactor or three reactors in series.

- Improve yields: In reaction systems such as $A \rightarrow B \rightarrow C$, 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.

- 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. So the heat of reaction is absorbed by the sensible heat required to rise the temperature of the excess material in the stream flowing through the reactor.

- Prevent side reactions: A large excess of one of the reactants 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.

- Control properties: In many polymerization rectors, 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

A plant's chemical species can be characterizes into three types: reactants, products, and inerts. A material balance for each of these components must be satisfied. This is typically mot a problem for products and inerts. However, the real problem usually arises when reactants are considered (because of recycle) and accounted 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, the loss of reactants exiting the process must be minimized since this represents a yield penalty. So reactants are prevented from leaving. Every mole of reactant fed to the process is consumed by the reactions.

This is an important concept and is generic to many chemical component balancing is not a problem because exit streams from the unit automatically adjust their flows and compositions. However, when units are connected 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.

3.3 Plantwide process control

Control analysis and control system design for chemical and petroleum processes have traditionally 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 processes contain a complex flowsheet with several recycle streams, energy integration, and many different unit operation. Essentially, the plantwide control problem is how to develop the control loops needed to operate an entire process and achieve its design objectives. Recycle streams and energy integration introduce a feedback of material and energy among units upstream and downstream. They also interconnect separate unit operations and create a pate for disturbance propagation. The presence of recycle streams profoundly alters that is not localized to an isolated part of the process.

Despite this process complexity, the unit operations approach to control system design has worked reasonably well. In the past, plants with recycle streams contained many surge tanks to buffer disturbances, to minimize interaction, and to isolate units in the sequence of material flow. This allowed each unit to be controlled individually. Prior to the 1970s, low energy costs meant little economic incentive for energy integration. However, there is growing pressure to reduce capital investment, working capital, and operating cost and to respond to safety and environmental concerns. This has prompted design engineers to start eliminating many surge tanks, increasing recycle streams, and introducing heat integration for both existing and new plants. 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 product yields, which reduce raw material costs, are achieved via lower reactant per-pass conversion and higher material recycle rates through the process. Better product quality, energy integration, and higher yields are all economically attractive in the steady-state flowsheet by they present significant challenges to smooth dynamic plant operation. Hence an effective control system regulating the entire plant operation and a 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 a 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 mot deal with component balances in the context of inventory control. By placing the priority on material balance over product quality controls, 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 conditions.

(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.

3.4 Basic Concepts of Plantwide Control

3.4.1. Buckley basics

Page Buckley (1964) was the first to suggest the idea of separating the plantwide control problem into two parts: material balance control and product quality control. He suggested looking first at the flow of material through the system. A logical arrangement of level and pressure control loops is established, using the flowrates of the liquid and gas process streams. No controller tuning or inventory sizing is done at this step. The idea is to establish the inventory control system by setting up this "hydraulic" control structure as the first step. He then proposed establishing the product-quality control loops by choosing appropriate manipulated variables. The time constants of the closed-loop product-quality loops are estimated. It is maked these as small as possible so that good, tight control is achieved, but stability constraints impose limitations on the achievable performance.

Then the inventory loops are revisited. The liquid holdups in surge volumes are calculated so that the time constants of the liquid level loops (using proportional-only controllers) are a factor of 10 larger than the product-quality tine constants. This separation in time constants permits independent tuning of the material-balance loops and the product-quality loops. Note that most level controllers should be proportional-only (P) to achieve flow smoothing.

3.4.2. Douglas doctrines

Jim Douglas(1988) has devised a hierarchical approach to the conceptual design of process flowsheets. Although he primarily considers the steady-state aspects of process design. He has developed several useful concepts that have control structure implications.Douglas points out that in the typical chemical plant the costs of raw materials and the value of the products are usually much greater than the costs of capital and energy. This leads to the two Douglas doctrines:

- Minimize losses of reactants and products.

- Maximize flowrates through gas recycle systems.

The first idea implies that need tight control of stream compositions exiting the process to avoid losses of reactants and products. The second 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 outweigh the additional energy cost of driving the recycle gas compressor.

The control structure implication is that do not attempt to regulate the gas recycle flow. It is simply maximized its flow. This removes one control degree of freedom and simplifies the control problem.

3.4.3. Downs drill

Jim Downs (1992) has insightfully pointed out 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..

But somehow these basics are often forgotten or overlooked in the complex and intricate project required to develop a steady-state design for a large chemical plant and specify its control structure. Often the design job is broken up into pieces. One person will design the reactor and its control system and someone else will design the separation section and its control system. The task sometimes falls through the cracks to ensure that these two sections operate effectively when coupled together. Thus it is important that perform the Downs drill.

All components (reactants, products, and inerts) must have a way to leave or be consumed within the process. The consideration of inerts is seldom overlooked. Heavy inerts can leave the system in the bottoms product from a distillation column. Light inerts can be purged from a gas recycle stream or from a partial condenser on a column. Intermediate inerts must also be removed in some way, for example in sidestream purges or separate distillation columns.

Most of the problems occur in the consideration of reactants, particularly when several chemical species are involved. All of the reactants fed into the system must either be consumed via reaction or leave the plant as impurities in the exiting streams. Since the plant usually want to minimize raw material costs and maintain high-purity products, most of the reactants 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. This is due to the fact that prevents reactants from leaving the process through composition controls in the separation section. Any imbalance in the number of moles of reactants involved in the reactions, no matter how slight, will result in the process gradually filling up with the reactant component that is in excess.

3.4.4. Luyben laws

Three law have been developed as a result of a number of case studies of many types of systems:

- A stream somewhere in all recycle loops should be flow controlled. This is to prevent the snowball effect.
- A fresh reactant feed stream cannot be flow-controlled unless there is essentially complete one-pass conversion of one of the reactants. This law applies to systems with reaction types. 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.
- If the final product from a process comes out 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 (Cantrell et al., 1995). Changes in feed folwrate or feed composition have less of a dynamic effect on

distillate composition then they do on bottoms composition if the feed is saturated liquid. The reverse is true if the feed is saturated vapor: bottoms is less affected than distillate. If our primary goal is to achieve tight product quality control, the basic column design should consider the dynamic implications of feed thermal conditions. Even if steady-state economics favor a liquid feed stream. The profitability of an operating plant with a product leaving the bottom of a column may be much better if the feed to the column is vaporized. This is another example of the potential conflict between steady-state economic design and dynamic controllability.

3.4.5. Richardson rule

Bob Richadson suggested 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 raft. 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.

3.4.6. Shinskey schemes

Greg Shinskey (1988) has 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 use should be made of ratio control, cascade control, override control, and valve-position (optimizing) control.

3.4 7. Tyreus tuning

One of the vital steps in developing a plantwide control system, once both the process and the control structure have been specified, is to determine the algorithm to be used for each controller (P, PI, or PID) and to tune each controller. The recommendation is using of P-only controllers for liquid levels (even in some liquid reactor applications). Tuning of a P controller is usually trivial: set the controller gain equal to 1.67. This will have the valve wide open when the level is at 80 percent and the valve shut when the level is at 20 percent (assuming the stream flowing out of the vessel is manipulated to control liquid level; if the level is controlled by the inflowing stream the action of the controller is reverse instead of direct).

For other control loops, The PI controllers is suggested to use. The relay-feedback test is a simple and fast way to obtain the ultimate gain (Ku) and ultimate period (Pu). Them either the Ziegler-Nichols settings (for very tight control with a closed-loop damping coefficient of about 0.1) or the Tyreus-Luyben (1992) settings (for more conservative loops where a closed-loop damping coefficient of 0.4 is more appropriate) can be used:

The use of PID controllers should be restricted to those loops where two criteria are both satisfied: the controlled variable should have a very large signal-to-noise ratio and tight dynamic control is really essential from a feedback control stability perspective.

3.5 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 constraints; liquid level and gas pressure inventories; makeup of reactants; component balances; and economic or process optimization.

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 setpoint. 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., minimize 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 from 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 flowrate of the limiting fresh reactant to the flowrate 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 setpoint. The setpoints 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 to selected.

When the setpoint 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 setpoint 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 the dynamic relationships between the controlled and manipulated variables feature small time constants and deadtimes 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 years, this shift in emphasis follows naturally.

The magnitudes of various flowrates 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 boilup) and base level is controlled with bottoms product flowrate. However, in columns with a large boilup ratio and small bottoms flowrate, these loops should be reversed because boilup 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 the remaining control degrees of freedom.

After satisfying all of the basic regulatory requirements, An additional degrees of freedom involving control valves that have not been used and setpoints in some controllers that can be adjusted. These can be used either to optimize steady-state economic process performance or to improve dynamic response.

Additional considerations

Certain quantitative measures from linear control theory may help at various steps to assess relationships between the controlled and manipulated variables. These include steady-state process gains, open-loop time constants, singular value decomposition, condition numbers, eigenvalue analysis for stability, etc. These techniques are described in detail in most process control textbooks. The plantwide control strategy should ultimately be tested on a nonlinear dynamic model that captures the essential process behavior

3.6 Plantwide control Problem

3.6.1.Units in Series

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. It is not had to worry about the issues discussed in the previous section and it can be 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 (flowrate 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 feed stream containing a mixture of chemical components A, B, and C is fed into a two-column distillation train. The relative volatilities are $\alpha_A > \alpha_B > \alpha_C$, and the "direct" (or "light-out-first") separation sequence is selected: A is taken out the top of the first column and B out the top of the second column.

Figure 3.1 (a) shows the situation where the fresh feed stream is flow-controlled into the process. The inventory loops (liquid levels) in each unit are controlled by manipulating flows leaving that unit. All disturbances propagate from unit to unit down the series configuration. The only disturbances that each unit sees are changes in its feed conditions.

Figure 3,1 (b) shows the on-demand situation where the flowrate of product C leaving the bottom of the second column is set by the requirements of a downstream unit. Now some of the inventory loops (the base of both columns) are controlled by manipulating the feed into each column.

When the units are arranged in series with no recycles, the plantwide control problem can be effectively broken up into the control of each individual unit operation. There is no recycle effect, no coupling, and no feedback of material from downstream to upstream units. The plant's dynamic behavior is governed by the individual unit operations and the only path for disturbance propagation is linear along the process.

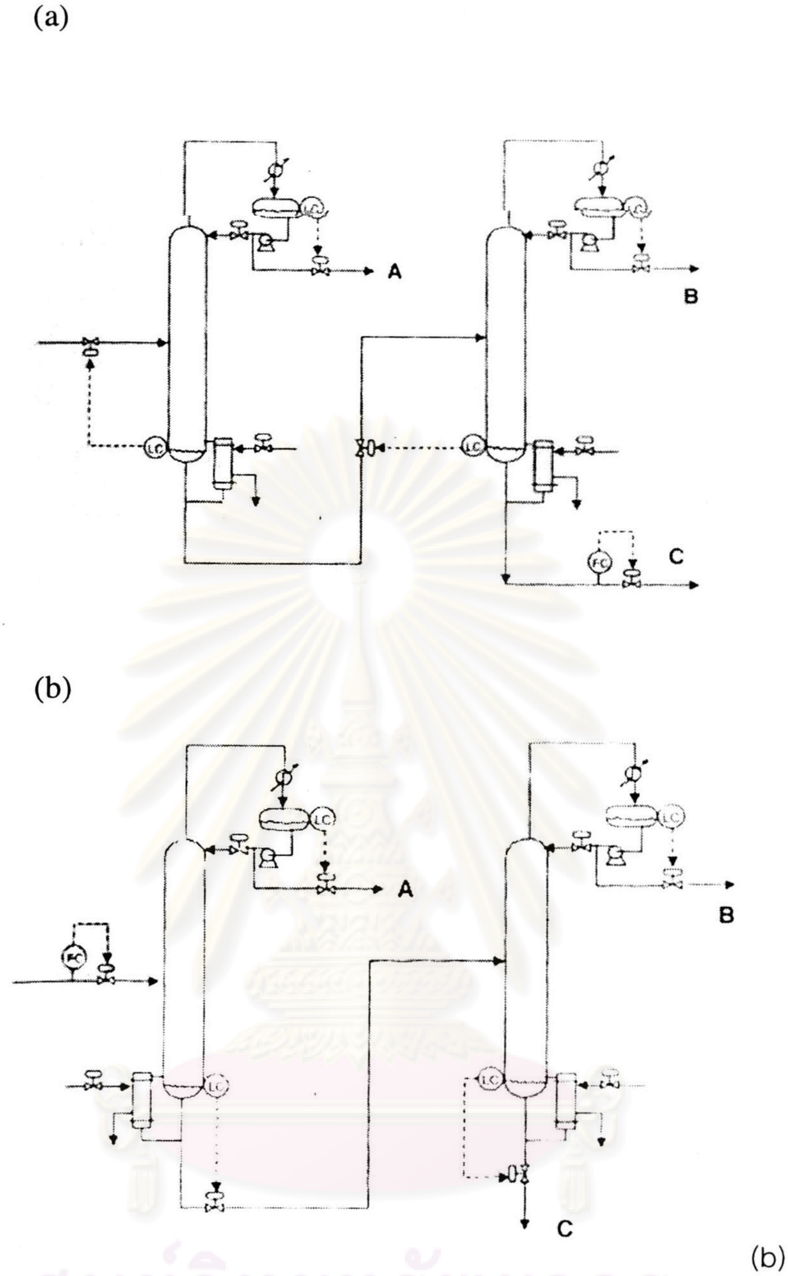

Figure 3.1 Unit in series. (a) Level control in direction of flow; (b) level control in direction opposite flow.

3.6.2 Effects of Recycle

Most real processes contain recycle streams. In this case the plantwide control problem becomes much more complex and its solution is not intuitively obvious. The presence of recycle streams profoundly alters the plant's dynamic and steady-state behavior. To gain an understanding of these effects, some very simple recycle systems are looked. The insight they are obtained from these idealized, simplistic systems can be extended to the complex flowsheets of typical chemical processes. First the groundwork must be laid and had some feel for the complexities and phenomena that recycle streams produce in a plant.

Two basic effects of recycle:

(1) Recycle has an impact on the dynamics of the process. The overall time constant can be much different than the sum of the time constants of the individual units.

(2) Recycle leads to the "snowball" effect. This has two manifestations, one steady state and one dynamic. A small change in throughput or feed composition can lead to a large change in steady-state recycle stream flowrates. These disturbances can lead to even larger dynamic changes in flows, which propagate around the recycle loop. Both effects have implications for the inventory control of components.

Snowball effects

Another interesting observation that has been made about recycle systems is their tendency to exhibit large variations in the magnitude of the recycle flows. Plant operators report extended periods of operation when very small recycle flows occur. It is often difficult to turn the equipment down to such low flowrates. Then, during other periods when feed conditions are not very different, recycle flowrates increase drastically, usually over a considerable period of time. Often the equipment cannot handle such a large load.

This high sensitivity of the recycle flowrates to small disturbances is called the snowball effect. It is important to note that this is not a dynamic effect; it is a steady-state phenomenon. But it does have dynamic implications for disturbance propagation and for inventory control. It has nothing to do with closed-loop stability. However, this does not imply that it is independent of the plant's control structure. On the contrary, the extent of the snowball effect is very strongly dependent upon the control structure used.

The large swings in recycle flowrates are undesirable in a plant because they can overload the capacity of the 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. As the example below illustrates and as more complex processes discussed in later chapters also show, a very plant wide control heuristic "A stream somewhere in each liquid recycle loop should be flow controlled".

Let us consider one of the simplest recycle processes imaginable: a continuous stirred tank reactor (CSTR) and a distillation column. As shown in Figure 3.2, a fresh reactant stream is fed into the reactor. Inside the reactor, a first-order isothermal irreversible reaction of component A to produce component B occurs A→ B. The specific reaction rate is k (h^{-1}) and the reactor holdup is V_R (moles). The fresh feed flowrate is F_0 (moles/h) and its composition is z_0 (mole fraction component A). The system is binary with only two components: reactant A and product B. The composition in the reactor is z (mole fraction A). The reactor effluent, with flowrate F (moles/h) is fed into a distillation column that separates unreacted A from product B.

The relative volatilities are such that A is more volatile than B, so the bottom from the column is the product stream. Its flowrate is B (moles/h) and its composition is x_B (mole fraction A). The amount of A impurity in this product stream is an important control objective and must be maintained at some specified level to satisfy the product quality requirements of the customer.

The overhead distillate stream from the column contains almost all of component A that leaves the reactor because of the purity specification on the bottoms stream. It is recycled back to the reactor at a flowrate D and with a composition x_D (mole fraction A). The column has trays and the feed tray is N_F (counting from the bottom). The reflux flowrate is R and the vapor boilup is V (moles/h).

The two alternative control structures for this process:

Conventional control structure As shown in Fig. 3.2, the following control loops are chosen:

1. Fresh feed flow is controlled.
2. Reactor level is controlled by manipulating reactor effluent flow.
3. Bottoms product purity is controlled by manipulating heat input to the reboiler.
4. Distillate purity is controlled by manipulating reflux flow. Note that dual composition control (controlling both distillate and bottoms purities) have been chosen to use in the distillation column, but there is no a priori reason for holding the composition of the recycle stream constant since it does not leave the process. It may be useful to control the composition of this recycle stream

for reactor yield purposes or for improved dynamic response. The "best" recycle purity levels in both the design and operation of the plant are been often free to find.

5. Reflux drum level is held by distillate flow (recycle).
6. Base level is held by bottoms flow.
7. Column pressure is controlled by manipulating coolant flowrate to the condenser.

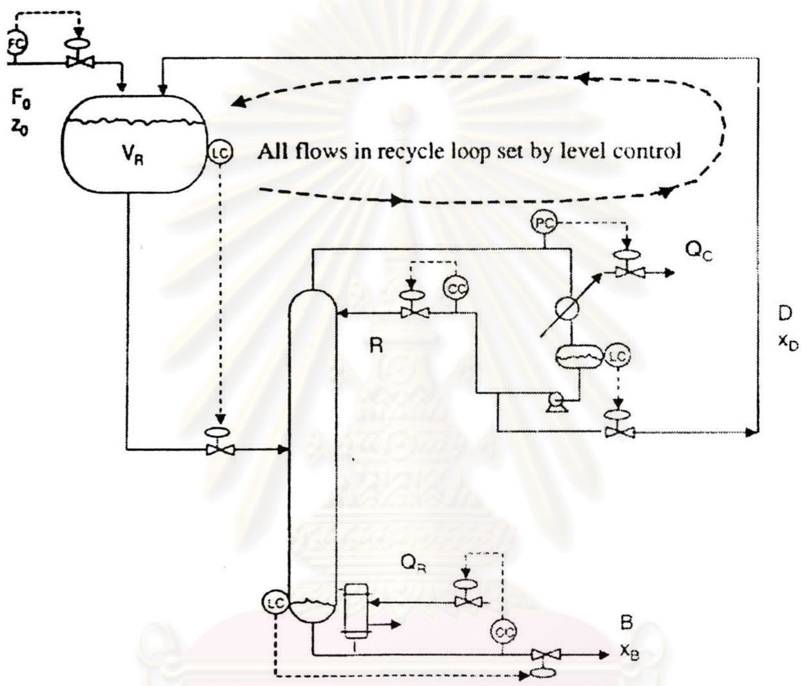

Figure 3.2 Conventional control structure with fixed reactor holdup

This control scheme is probably what most engineers would devise if given the problem of designing a control structure for this simple plant. Our tendency is to start with setting the flow of the fresh reactant feed stream as the means to regulate plant production rate, and then work downstream from there as if looking at a steady-state flowsheet and simply connect the recycle stream back to the reactor based upon a standard control strategy for the column.

However, this strategy is no flow controller anywhere in the recycle loop. The flows around the loop are set based upon level control in the reactor and reflux drum. This control structure is expected to find that exhibiting the snowball effect. By writing the various overall steady-state mass and component balances around the whole process and around the reactor and column, the flow of the recycle stream can be calculated at steady state for any given fresh reactant feed flow and composition.

With the control structure in Fig. 3.2 and the base-case fresh feed flow and composition, the recycle flowrate is normally 260.5 moles/h. However, the recycle flow must decrease to 205 moles/h when the fresh feed composition is 0.80 mole fraction A. It must increase to 330 moles/h when the fresh feed composition changes to pure A. Thus a 25 percent change in the disturbance (fresh feed composition) results in a 60 percent change in recycle flow. With this same control structure and the base-case fresh reactant feed composition, the recycle flow drops to 187 moles/h if the fresh feed flow changes to 215 moles/h. It must increase to 362 moles/h when the fresh feed flowrate is changed to 265 moles/h. Thus a 23 percent change in fresh feed flowrate results in a 94 percent change in recycle flowrate. These snowball effects are typical for many recycle systems when control structures such as that shown in Figure 3.2 are used and there is no flow controller somewhere in the recycle loop.

Variable reactor holdup structure An alternative control structure is shown in Figure 3.3. This strategy differs from the previous one in two simple but important ways.

1. Reactor effluent flow is controlled.
2. Reactor holdup is controlled by manipulating the fresh reactant feed flowrate.

All other control loops are the same. The production rate cannot change directly by manipulating the fresh feed flow, because it is used to control reactor level. However, The plant throughput can achieved indirectly in this scheme by changing the setpoint of the reactor level controller. Using the same numerical case considered previously, the recycle flowrate does not change at all when the fresh feed composition changes. To alter production rate from 215 moles/h to 265 moles/h (a 23 percent change), the reactor holdup must be changed from 1030 moles/h to 1520 moles/h (a 48 percent change), Recycle flow also changes, but only from 285 to 235 moles/h. This is an 18 percent change in recycle flow compared with 94 percent in the alternative strategy.

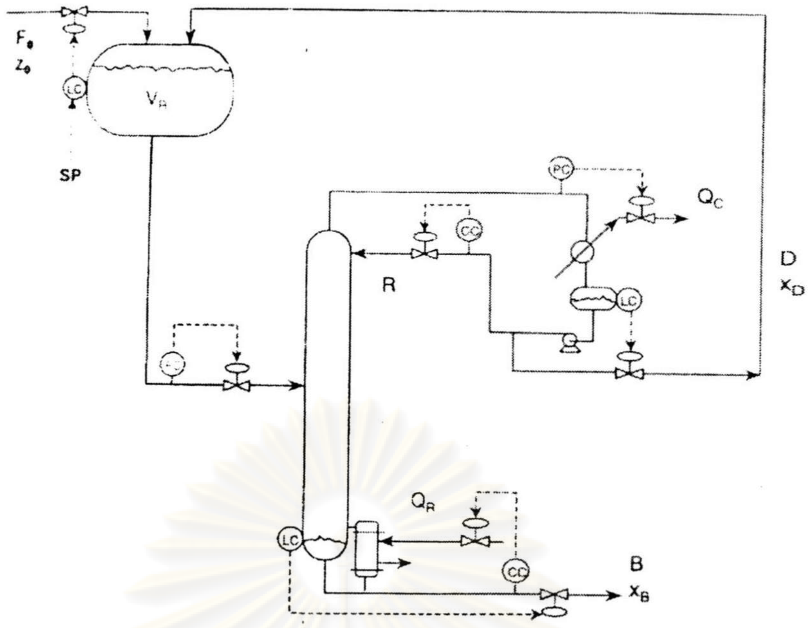

Figure 3.3 Control structure with variable reactor holdup

3.6.3 Reaction/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 constant), and reactant composition (mole fraction A) R = $V_R\, kz$. 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. For example, to increase production rate of component B by 20 percent, the overall reaction rate must increase by 20 percent Since both reactor temperature and reactor holdup V_R are held constant, reactor composition z must increase 20 percent. 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 flowrates.

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.

If the tuning of the reactor level controller in the conventional structure (Fig. 3.2) is modified from normal PI to P only, then changes in production rate also produce changes in reactor holdup. This tends to compensate somewhat for the required changes in overall reaction rate and lessens the impact on the separation section. So both control system structure and the algorithm used in the inventory controller of the reactor affect the amount of this snowball phenomenon.

This example has a liquid-phase reactor, where volume can potentially be varied. If the reactor were vapor phase, reactor volume would be fixed. However, an additional degree of freedom are had and could vary reactor pressure to affect reaction rate.

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 products in reversible reactions), 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 have fewer available manipulators than variables that would like to control. The setpoints of the partial control loops are then manipulated to hold the important economic objectives in the desired ranges.

The plantwide control implication of this idea is that production rate changes should preferentially be achieved by modifying the setpoint of a partial control loop in the reaction section. This means that the separation section will not be significantly disturbed. Using the control structure in Fig. 3.2, changes in production rate require large changes in reactor composition, which disturb the column. Using the control structure shown in Fig. 3.3, changes in production rate are achieved by altering the setpoint of a controlled dominant variable, reactor holdup, with only small changes in reactor composition. This means that the column is not disturbed as much as with the alternative control scheme.

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.