ПОШАГОВАЯ МЕТОДИКА ПРОЕКТИРОВАНИЯ УПРАВЛЕНИЯ ТЕХНОЛОГИЧЕСКИМ ПРОЦЕССОМ В МАСШТАБЕ ПРЕДПРИЯТИЯ

STEPS OF PLANTWIDE PROCESS CONTROL DESIGN PROCEDURE

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Аннотация

В данной статье составлена пошаговая методика проектирования управления технологическим процессом в масштабе предприятия. Вы узнаете, как наиболее эффективно управлять технологическим процессом на производстве. Самым важным аспект проблемы является, стационарное проектирование и задачи динамического управления процессом, потому что различные цели управления приводят к различным структурам управления. Есть старая персидская поговорка "Если ты не знаешь, куда идешь, любая дорога приведет тебя туда!". Это, безусловно, верно для контроля на всей территории завода.

Annotation

This article presents a step-by-step method for designing process control at the enterprise scale. You will learn how to manage the production process most

effectively. The most important aspect of the problem is stationary design and dynamic process management tasks, because different management goals lead to different management structures. There is an old Persian saying "If you don't know where you are going, any road will lead you there!". This is certainly true for controls throughout the plant.

Ключевые слова: Процесс, производительность, установка, проект, шаг.

Keywords: Process, performance, installation, project, step.

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. There is an old Persian saying “If you don’t know where you are going, any road will get you there!” This is certainly true in plantwide control. The “best” control structure for a plant depends upon the design and control criteria established.

These objectives include reactor and separation yields, product quality specifications, product grades and demand determination, environmental restrictions, and the range of safe operating 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, and (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).

During the course of the subsequent steps, we may find that we lack suitable manipulators to achieve the desired economic control objectives. Then we must change the process design to obtain additional handles. For example, we may need to add bypass lines around heat exchangers and include auxiliary heat exchangers.

Step 3: Establish energy management system

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

We use the term energy management to describe two functions: (1) We must 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) If heat integration does occur between process streams, then the second function of energy management is 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). More details of reactor control are discussed in Chap. 4.

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.

For example, suppose the temperature of a stream being fed to a distillation column is controlled by manipulating steam flowrate to a feed preheater. And suppose the stream leaving the preheater is partially vaporized. Small changes in composition can result in very large changes in the fraction of the stream that is vaporized (for the same pressure and temperature). The resulting variations in the liquid and vapor rates in the distillation column can produce severe upsets.

Heat integration of a distillation column with other columns or with reactors is widely used in chemical plants to reduce energy consumption. While these designs look great in terms of steady-state economics, they can lead to complex dynamic behavior and poor performance due to recycling of disturbances. If not already included in the design, trim heaters/coolers or heat exchanger bypass lines must be added to prevent this. Energy disturbances should be transferred to the plant utility system whenever possible to remove this source of variability from the process units. Chapter 5 deals with heat exchanger systems.

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, we must increase overall reaction rates. 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 we select 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, we can increase temperature 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, we can decrease reactor temperature 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, we must find another dominant variable, 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 we identify the dominant variables, we must also identify the manipulators (control valves) 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 we choose, we would like it to provide smooth and stable production rate transitions and to reject disturbances. We often want to select 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.

When the setpoint of a dominant variable is used to establish plant production rate, the control strategy must ensure that the right 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. We must keep these ideas in mind when we reach Steps 6 and 7.

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 we must somehow adjust the setpoint of the dominant variable on the basis of the production rate that has been specified externally. We must balance production rate with what has been specified externally. 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. This concept was nicely illustrated by the control strategy in Fig. 2.16. In that scheme we fixed externally the flow of fresh reactant A feed. Also, we used reactor residence time (via the effluent flowrate) as the controlled dominant variable. Feedback information (internal reactant composition information) is provided to this controller by the ratio of the two recycle stream flows.

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.

We want tight control of these important quantities for economic and operational reasons. Hence we should select manipulated variables such that the dynamic relationships between the controlled and manipulated variables feature small time constants and deadtimes and large steady-state gains. The former gives small closed-loop time constants and the latter prevents problems with the rangeability 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 in 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 steam 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: Fix a flow in every recycle loop and control inventories (pressures and levels)

Fix a flow in every recycle loop and then select the best manipulated variables to control inventories.

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 flows that can occur if all flows in the recycle loop are controlled by levels, as illustrated by the simple process examples in Chap. 2. Steady-state and dynamic benefits result from this flow control strategy. From a steady-state viewpoint, the plant's separation section is not forced to operate at significantly different load conditions, which could lead to turndown or flooding problems.

From a dynamic viewpoint, whenever all flows in a recycle loop are set by level controllers, wide dynamic excursions can occur in these flows because the total system inventory is not regulated. The control system is attempting to control the inventory in each individual vessel by changing the flowrate to its downstream neighbor. In a recycle loop, all level controllers see load disturbances coming from the upstream unit. This causes the flowrate disturbances to propagate around the recycle loop. Thus any disturbance that tends to increase the total inventory in the process (such as an increase in the fresh feed flowrate) will produce large increases in all flowrates around the recycle loop.

Fixing a flowrate in a recycle stream does not conflict with our discussion of picking a dominant reactor variable for production rate control in Step 4. Flow

controlling a stream somewhere in all recycle loops is an important simple part of any plantwide control strategy.

Gas recycle loops are normally set at maximum circulation rate, as limited by compressor capacity, to achieve maximum yields (Douglas doctrine).

Once we have fixed a flow in each recycle loop, we then determine what valve should be used to control each inventory variable. This is the material balance step in the Buckley procedure. Inventories include all liquid levels (except for surge volume in certain liquid recycle streams) and gas pressures. An inventory variable should typically be controlled with the manipulated variable that has the largest effect on it within that unit (Richardson rule). Because we have fixed a flow in each recycle loop, our choice of available valves has been reduced for inventory control in some units. Sometimes this actually eliminates the obvious choice for inventory control for that unit. This constraint forces us to look outside the immediate vicinity of the holdup we are considering.

For example, suppose that the distillate flowrate from a distillation column is large compared to the reflux. We normally would use distillate to control level in the reflux drum. But suppose the distillate recycles back to the reactor and so we want to control its flow. What manipulator should we use to control reflux drum level? We could potentially use condenser cooling rate or reboiler heat input. Either choice would have implications on the control strategy for the column, which would ripple through the control strategy for the rest of the plant. This would lead to control schemes that would never be considered if one looked only at the unit operations in isolation.

Inventory may also be controlled with fresh reactant makeup streams as discussed in Step 4. Liquid fresh feed streams may be added to a location where level reflects the amount of that component in the process. Gas fresh feed streams may be added to a location where pressure reflects the amount of that material in the process.

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 flowrate disturbances to the downstream separation system. There is nothing necessarily sacred about holding reactor level constant.

Step 7: Check component balances

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

Component balances can often be quite subtle, but they are particularly important in processes with recycle streams because of their integrating effect. 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 reaction rate in the reactor. The buildup of chemical components in recycle streams must be prevented by keeping track of chemical component inventories (reactants, products, and inerts) inside the system.

We must identify the specific mechanism or control loop to guarantee that there will be no uncontrollable buildup of any chemical component within the process (Downs drill).

What are the methods or loops to ensure that the overall component balances for all chemical species are satisfied at steady state? We can limit their intake, control their reaction, or adjust their outflow from the process.

As we noted in Chap. 2, we can characterize a plant's chemical components into reactants, products, and inerts. We don't want reactant components to leave in the product streams because of the yield loss and the desired product purity specifications. Hence we are limited to the use of two methods: consuming the reactants by reaction or adjusting their fresh feed flow. Product and inert components all must have an exit path from the system. In many systems inerts are removed by purging off a small fraction of the recycle stream. The purge rate is adjusted to control the inert composition in the recycle stream so that an economic balance is maintained between capital and operating costs.

We recommend making a Downs drill table that lists each chemical component, its input, its generation or consumption, and its output. This table should

specify how the control system will detect an imbalance in chemical components and what specific action it will take if an imbalance is detected.

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 individual chemical units (Shinskey, 1988), For example, a tubular reactor usually requires control of inlet temperature. High-temperature endothermic reactions typically have a control system to adjust the fuel flowrate to a furnace supplying energy to the reactor. Crystallizers require manipulation of refrigeration load to control temperature. Oxygen concentration 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 to the gas feed. We deal with the control of various unit operations in Chaps. 4 through 7.

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, we usually have additional degrees of freedom involving control valves that have not been used and setpoints in some controllers that can be adjusted. These can be utilized either to optimize steady-state economic process performance (e.g., minimize energy, maximize selectivity) or to improve dynamic response.

For example, suppose an exothermic chemical reactor may be cooled with both jacket cooling water and brine (refrigeration) to a reflux condenser. For fast reactor temperature control, manipulation of brine is significantly better than cooling water. However, the utility cost of brine is much higher than cooling water. Hence we would like the control system to provide tight reactor temperature control while minimizing brine usage. This can be achieved with a valve position control strategy. Reactor temperature is controlled by manipulating brine, A valve position controller

looks at the position of the brine control valve and slowly adjusts jacket cooling water flow to keep the brine valve approximately 10 to 20 percent open under steady-state operation (Fig. 3.2)

Список литературы

1. Проектирование и управление реактивной дистилляцией. Автор William L. Luyben and Cheng-Ching Yu Copyright © 2008 John Wiloy & Sons, Inc.
2. Проектирование Химических Процессов: Компьютерные Тематические Исследования. Alexandre C. Dimian and Constin Sorin Bildea Copyright © 208 WILEY-VCH Verlag GmbH & Co. KGaA, Wcinhcim.
3. Westerberg, A. W., O. M. Wahnshaft, Synthesis of distillation based separation processes, in ADV. Chem. Морская щука., Том. 23, Process Synthesis, Acdemic Press, 1996

Bibliography:

1. Reactive Distillation Design and Control. By William L. Luyben and Cheng-Ching Yu Copyright © 2008 John Wiloy & Sons, Inc.
2. Chemical Process Design: Computer-Aided Case Studies. Alexandre C. Dimian and Constin Sorin Bildea Copyright © 208 WILEY-VCH Verlag GmbH & Co. KGaA, Wcinhcim.
3. Westerberg, A.W., O.M. Wahnshaft, Synthesis of distillation based separation processes, in Adv. Chem. Ling., Vol. 23, Process Synthesis, Acdemic Press, 1996