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Process Dynamics and Process Control Overview

CHAPTER CONTENTS

1.1 Introduction
1.2 The Role of Process Control
1.3 Tag Naming Conventions
1.3.1 Tag Naming in Process Overview/Control Strategy Drawings
1.3.2 Process Overview/Control Strategy Drawing Conventions
1.4 Control Loop Essentials
1.4.1 The Feedback Control Loop
1.4.2 Control Loop Hardware
1.4.3 Control Loop Examples
1.4.4 Controller Modes
1.4.5 Open-loop Stable and Unstable Control Loops
1.5 Process Dynamics and Dynamic Responses
1.5.1 Representation of Process Dynamics: Process Response Curves
1.5.2 First-Order Linear Continuous Model
1.5.3 Transfer Function Representation
1.5.4 First-order Linear Discrete Model
1.6 Plant Testing
1.6.1 Objectives
1.6.2 Work Process
1.6.3 In Process Control
1.7 Classification of Process Control Strategies
1.7.1 Control Strategy Hierarchy: Relative Costs, Resources, and Credits
1.7.2 Safety Protection
1.7.3 Basic Regulation: Basic Regulatory Control
1.7.4 Advanced Regulation: Advanced Regulatory Control
1.7.5 Supervisory Control: Conventional Advanced Control Applications
1.7.6 MVCC and RTO: Advanced Control Applications
1.8 Benefits of Control Applications
References

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While process unit design is based on steady-state principles and relationships, the process unit's operating environment is dynamic, characterized by changing feedstocks, unexpected upsets, equipment failures, changing operating directives, and the need to satisfy constraints, some of which are time-dependent. Maintaining plant operations at desired targets, meeting changing operating objectives, eliminating the effects of process upsets, and maintaining operations within and at constraint limits is the domain of process control. Process dynamics refers to the dynamic relationships among process variables (PVs) that determine the process' time-dependent response to routine operator adjustments, unexpected upsets, hardware failures, and transitions

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to new operations. Process dynamics understanding plays an essential role in the design and implementation of process control strategies.

1.1 Introduction

This chapter begins with an explanation of how important components of the feedback control loop – measured inputs, calculated inputs, control loops, and calculated control outputs – are referenced in a modern Digital Control System (DCS). All such entities are assigned a *name,* made up of alphanumeric characters, and the name assigned to each entity is called its *tag name*. Throughout the book, standardized drawings are used to present process unit configurations and control strategy cascades, with control relevant components identified by their tag names. This corresponds to real-world project documentation used by operating sites, engineering, procurement and construction (EPC) contractors and DCS vendors. Therefore, to serve as future references, the tag naming and drawing conventions are described and summarized.

The control loop is the basic unit of process control. Generic feedback control loops are presented to illustrate the relationships among a control loop's control variable (CV), manipulated variable (MV), and disturbance variable (DV). Several control loop examples (flow, pressure, and level controllers) are presented, and the important role of controller mode is explained.

A dependent variable's (e.g., a CV) time-domain dynamic response to a change in an independent variable (e.g., an MV or DV) has characteristics (steady-state gain, K_P, time-to-steady-state [TTSS], time constant, τ, and deadtime, θ) that are relevant to process understanding, control strategy design, and future control performance. A generic dependent variable's dynamic response to a step change in an independent variable is used to illustrate the physical significance and importance of K_P, TTSS, τ, and θ, and simple graphical procedures are presented to illustrate how these model parameters are estimated. These parameters are used throughout this chapter to explain the fundamentals of process dynamics and in subsequent chapters to guide control strategy designs.

In a modern DCS, data is sampled at a fixed frequency and process information is known only at discrete times, making classical time-domain approaches invalid. Therefore, the discrete equivalent of the time domain's response curve – a recursive difference equation that contains dependent and independent variable values at specific time steps (but with the same model parameters) is presented and discussed. This discussion facilitates the understanding of the discrete PID (proportional, integral, derivative) control algorithm and its basic functionality presented in Section 2.4 and the discrete linear exponential filter presented in Section 2.5.

Plant tests are performed to increase understanding of dynamic relationships among process variables (PV) and/or to identify a specific response's dynamic parameters. Although simple graphical procedures are presented to illustrate how K_P, TTSS, τ, and θ are estimated, commercially available statistical curve-fitting tools are far more effective compared to graphical methods at obtaining accurate model parameter values. These methods are described qualitatively, with the important role of the residual (the difference between predicted and measured PV values at each time step)[1] explained.

Process control strategies are categorized (via five "Levels") based on control need – safety protection is the most important need (Level 1), followed by basic regulatory controls (Level 2), advanced regulatory controls (Level 3), conventional supervisory controls (Level 4), and advanced control applications, e.g., dynamic matrix control (DMC), RTO (Level 5). Although the focus of this book is the design of control strategies at Levels 3 and 4, the control functionality each level provides is summarized and illustrated with industrially important applications.

1 In the discussion of model-based control in Sections 2.7, model error and prediction error are related to residuals.

Finally, how control strategies improve process control performance and capture credits is described – by reducing a CV's variability, operations can be moved closer to a product specification or constraint limit. However, unless operations are actually moved closer to the specification or constraint limit by a CV set point change, the bulk of the potential benefits are lost.

1.2 The Role of Process Control

The non-steady-state nature of refinery and chemical plant operations makes process control an essential requirement for successful plant operations. Process units are designed to operate at a set of desirable *steady state* operating points *within* an expected operating region or "window." Steady-state material and energy balances, physical models of unit operations, thermodynamic relationships, and design correlations are used in unit designs. Typically, multiple cases are considered in the unit's design, e.g., winter and summer operations, catalyst start-of-run/end-of-run status, various feed rates and feed types, and several desired products and product qualities. Various process, equipment, thermodynamic, physical and/or chemical equilibrium, safety, and environmental constraints bound unit operations to a set of feasible states – the operating window. The first task for well-designed and well-tuned process control strategies is to hold plant operations at desired operating points within the operating window despite unexpected process upsets/disturbances, e.g., feed composition and temperature changes, fuel gas composition changes, instrument and other hardware failures, weather upsets, process equipment failures, and operator errors. Profitable operations typically include many PVs at high or low constraint limits. Therefore, process control strategies, designed to meet control and operating objectives, must also safely respect multiple process constraints.

With plant operations at a desired operating point within the operating window, operators will frequently make *routine changes* in some units. Examples of common, routine changes are (i) relieve bottlenecks (e.g., full open valve, condenser cooling limitation), (ii) change unit feed rates, (iii) increase reactor feed preheat to increase feed temperature or increase reactor treat gas/feed ratio to increase reactor conversion, and (iv) change a light-ends tower's reboiler steam flow rate or its reflux flow rate to help bring product quality back to target. The plant's dynamic responses to typical process adjustments depend directly upon the performance of the plant's suite of underlying process control strategies. Thus, process control strategies must effectively manage routine operator adjustments as process conditions change and/or as new limiting constraints become active.

Profit varies within the operating window. Therefore, as business drivers change, there are incentives to identify the process conditions within the operating window that align the plant's operations with the new economics. The business team, which includes site planners and coordinators, typically employs a variety of offline tools to analyze the impact of business drivers on plant operations. The business team, with input from site operations, maintenance, and technical personnel, develops an operating plan. The plan contains a set of operating objectives and the new set of process conditions that will achieve the new operating objectives. Important information specified includes new flow, temperature and pressure targets, desired products and their qualities, potential limiting constraints, PV high/low limits, feed and product tank IDs, and equipment to place in or take out of service. Business plans are typically provided weekly or more frequently if needed to account for unforeseen economic changes.

Process control is essential for operators to guide the plant's transition to the new, desired operation in a smooth and stable manner in a reasonable timeframe while honoring process constraints. Operators make the necessary process adjustments (e.g., changes to controller set points and PV

high/low constraint limits) via the DCS operator's console; some manual changes to field hardware may also be needed. The plant's transition to the new operating point is monitored using the DCS' operator interface.

In summary, the primary role of process control is to accomplish the following tasks:

- Hold plant operations at the current, desired operating point despite unexpected upsets/disturbances.
- Provide smooth, fast, and safe transitions from one operation to another when (i) routine operator adjustments are made as some process conditions and/or limiting constraints change and (ii) the business team issues a new operating plan that must be implemented.
- Keep plant operations on the safe side of limiting constraints.

1.3 Tag Naming Conventions

Section 1.4 below will reveal that measured inputs, calculated inputs, control loops, and calculated control outputs are all important components of the feedback control loop. These feedback loop components are all assigned names in the DCS; the name assigned to each is called its tag name. As used in this book, tag names have the following general format[2]:

SRCxyz

where

SRC – service code
xyz – numerical identifier

The service code is either two or sometimes three alphabetic characters. The first character indicates the tag's PV input source type – for example, *F* means flow and *Q* means heat duty. The service code's second character indicates the tag's function – for example, *C* means controller. The third character in the service code is optional and is used infrequently. But when it is used, it indicates the special task associated with the tag – for example, *V* means validation. Table 1.1 lists the characters used and what they represent. Here are some examples:

FI89 – Measured flow input
LI9 – Measured level input
PC134 – Pressure control loop
TC387 – Temperature control loop
TV387[3] – TC387 sends its output (% range) to this valve
AIV243 – Analyzer measurement (AI243) validation
FCG114 – Feed flow FC114 lead/lag dynamic compensation (feedforward control)

All tags have *tag attributes*, specific properties or values that have assigned names and mnemonics in the DCS. Example properties are engineering units (EUs) and engineering unit range (EUHI, EULO), and example values are measured or calculated controller inputs (PV) and controller set points (SPs). In this book, when referring to a tag and its attributes, the tag name and tag attribute mnemonic are combined with a blank character between them. For example, FC119 SP is the reference for flow controller FC119's SP, and FC119 PV is the reference for FC119's PV input.

2 In a plant, tag names start with a unit code, typically two alphabetic (or sometimes numerical) characters – for example PS for the atmospheric pipestill unit. Because tag name duplication and tag name assignment to an operator's console are not issues here, unit codes are not needed and they are omitted.
3 As explained in Section 2.4.3, see Figure 2.14, TV387 would be a name assigned to the analog output (AO) block (i.e., the valve) that receives TC387's control output (in % of range).

Table 1.1 Service Codes – First, Second, and Third Characters

First Character		Second Character		Third Character	
A	Analyzer	C	Control	G	Dynamic Compensation
F	Flow	I	Indicator	V	Validation
L	Level	V	Valve	Z	Calculation
P	Pressure	—	—	—	—
Q	Heat Duty	—	—	—	—
R	Ratio	—	—	—	—
S	Speed	—	—	—	—
T	Temperature	—	—	—	—

1.3.1 Tag Naming in Process Overview/Control Strategy Drawings

The focal point of almost all control strategy discussions throughout the book are two drawings. The first drawing provides a process configuration overview and is the basis for the process description. The second drawing presents the control strategy's cascade structure and is the basis for the strategy's functional description. For future reference, the following "rules" highlight how the non-instrument tags, the application tags, in a control strategy's cascade structure are named. They are presented in "example form" for the most common situations:

1. An application tag doing temperature control with an input source from a (measured) high-level analog input (4–20 mA) temperature indicator T**I**311 is named T**C**311 (I → C, with the same service code first character, and the same numerical identifier as PV-input source).
2. An application tag doing flow control with an input source from a (measured) flow indicator F**I**103 is named F**C**103 (I → C, with the same service code first character, and the same numerical identifier as PV-input source).
3. An application tag validating an online analyzer measurement AI186 is named AI**V**186 (3rd character **V** added to service code, but the same first and second service code characters, and the same numerical identifier as measurement input source).
4. An application tag validating a DTI point TI343 is named TI**V**343 (3rd character **V** added to service code, same first and second service code characters, and the same numerical identifier as measurement input source).
5. An application tag providing dynamic compensation for flow measurement FI45 is named FI**G**45 (3rd character **G** added to service code, same first and second service code characters, and the same numerical identifier as measurement input source).
6. An application tag providing dynamic compensation for a flow measurement controlled by flow controller FC127 is named FC**G**127 (3rd character **G** added to service code, same first and second service code characters, and the same numerical identifier as measurement input source).
7. An application's ratio controller, outputting to flow controller FC243, is named **RC**243 (new service code **RC**, same numerical identifier as secondary).
8. An application's ratio controller, outputting to heat duty controller QC300, is named **RC**300 (new service code **RC**, same numerical identifier as secondary).
9. An application's reboiler heat duty controller, outputting to flow controller FC38, is named **QC**38 (new service code **QC**, same numerical identifier as secondary).
10. An application's reboiler heat duty controller, outputting directly to control valve TV281, is named **QC**281 (new service code QC, same numerical identifier as secondary).

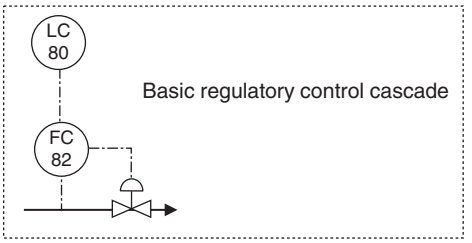
1.3.2 Process Overview/Control Strategy Drawing Conventions

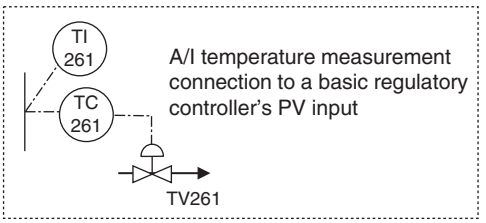
The drawing conventions used for both the process overview and the control strategy's cascade structure drawings are summarized in Table 1.2. Notice that many of the tag naming rules presented in Section 1.3.1 are illustrated in Table 1.2.

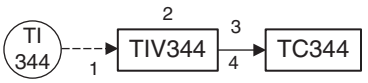
The tag naming and drawing conventions summarized in Tables 1.1 and 1.2 are used initially in Section 1.4.3, and then in most subsequent sections of the book.

Table 1.2 Process Overview and Control Strategy Drawing Conventions

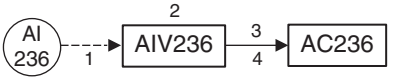






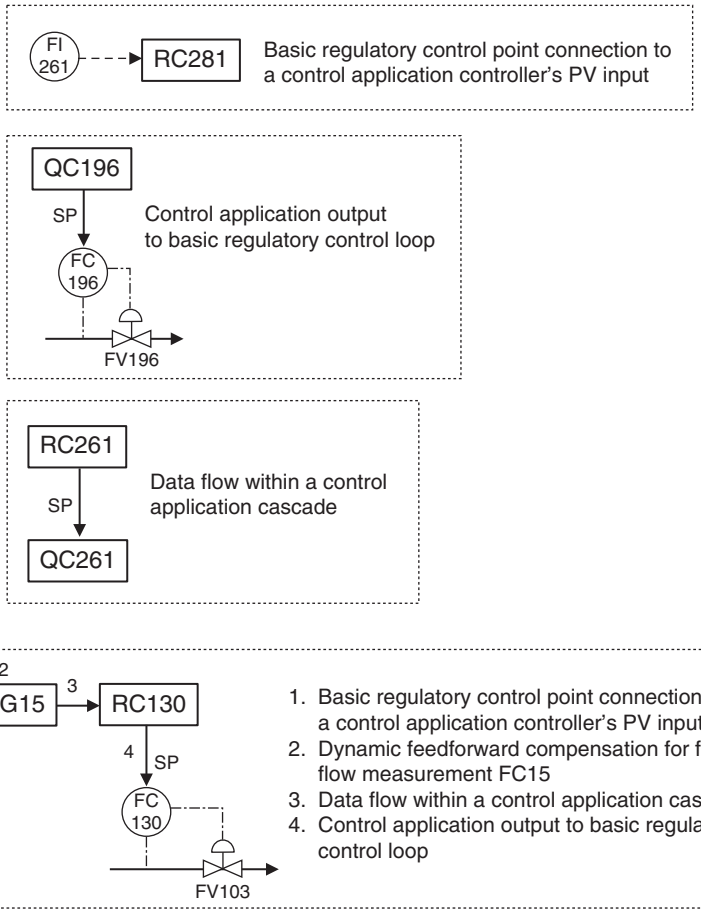


1. Basic regulatory control point connection to a control application controller's PV input
2. DTI temperature measurement validation
3. Data flow within a control application cascade
4. Validated DTI temperature connection to control application controller's PV input



1. Basic regulatory control point connection to a control application controller's PV input
2. Analyzer measurement validation
3. Data flow within a control application cascade
4. Validated analyzer measurement connection to control application controller's PV input

Table 1.2 (Continued)



1.4 Control Loop Essentials

A PV describes a process condition; if it is maintained at a desired target within or at the boundary of an operating region via closed-loop control, it is called a CV. A CV can be a measured PV, a PV calculated from measured PVs, a lab value, or a final control element, e.g., a valve position. Most measured CVs are temperatures, pressures, flows, levels, and compositions. Typical calculated CVs are tower internal reflux,[4] tower reboiler duty,[5] and fired heater duty.[6] CVs are *dependent variables* whose measured or calculated values are determined by the process configuration and the current values of *other PVs*.

A PV value calculated by a controller to achieve a control objective, e.g., maintain a CV at a desired target (the CV's *set point*), is called an MV. MVs change a valve position or other actuator

4 Defined and discussed in Chapter 3, Section 3.7.2.
5 Discussed in Chapter 3, Section 3.7.1.
6 Discussed in Chapter 10, Section 10.4.6.

(e.g., fired heater stack damper), or in a cascade, an MV changes the SP of another controller. Ultimately, it is a valve position or actuator action that changes process conditions; in a cascade, the *ultimate secondary* outputs to the final control element. MVs are *independent variables* whose values are set by either controller adjustments or operator entries and can be maintained at desired targets as process conditions change. CV values vary as controller adjustments and operator entries change MV values.

A PV that upsets or disturbs a control loop is called a DV. DVs are *independent* variables, measured or unmeasured, that affect the process and as the process responds to the DV's change, one or more CVs are affected. For example, a light-ends tower's feed rate is frequently set by upstream process conditions, but when it changes, it can have a significant effect on the tower's overhead and/or bottoms product's quality.[7] Also, upsets in any unit providing off gas to the fuel gas header can cause rapid fuel gas heating value changes which can have an important impact on fired heater coil outlet temperature controls.[8] To correct for a disturbance, and return their CV to SP, the affected feedback control loop responds by making MV adjustments.

1.4.1 The Feedback Control Loop

The *control loop* is the basic unit of process control. A generic feedback control loop that illustrates the relationships among a CV, MV, and DV appears in Figure 1.1. In Figure 1.1, the CV is a measured PV which is the feedback controller's PV input. Notice that the measurement hardware is an element of the control loop. In this book, the controller is almost always a PID controller, and its control output is calculated based on the difference between the CV value and its SP, the *controller error*. Even though error will always be understood to be the difference between a CV value and its SP, we will normally refer to error more generically as the difference between PV and SP, e.g., $Error \equiv PV–SP$. The controller's calculated output determines the new value of the MV. Notice the actuator hardware is also an element of the control loop. The CV responds to new MV values, and

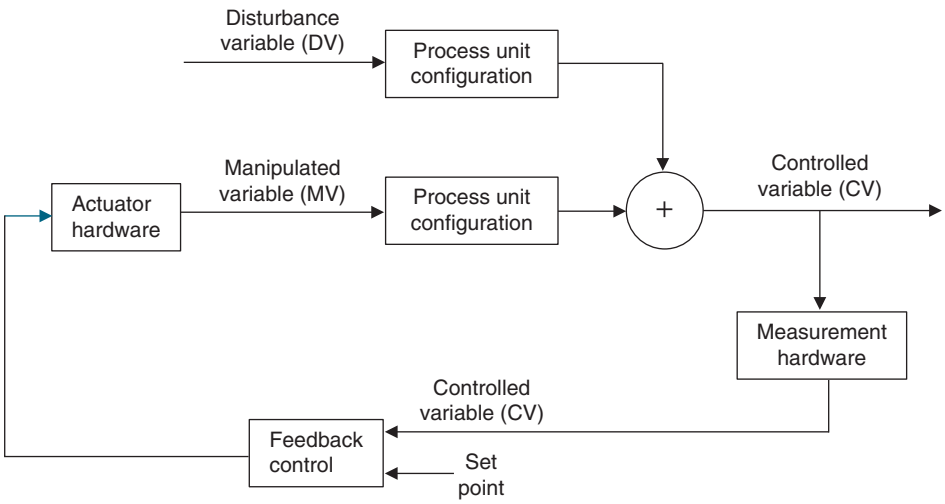


Figure 1.1 Feedback Control Loop Structure

7 Light-ends tower controls are presented and discussed in Chapter 12 (Tower Product Quality Control).
8 Fired heater firing controls are presented and discussed in Chapter 10, Section 10.4.

new CV values are measured and become the controller input. This sequence continues until the CV approaches its SP, error approaches zero, and there are no further MV changes.

Nonzero MV adjustments occur whenever the controller error (PV–SP) is nonzero, e.g., when either an operator or another controller in a cascade changes the SP, or a process upset occurs that affects the CV (the error's PV input). With the CV initially at its SP, when the SP is changed, controller error immediately jumps, and the controller calculates a new MV value. The CV responds to this new MV value, and a new CV value is measured and becomes the controller's input. This sequence continues until the CV approaches the new SP, error approaches zero, and there are no further MV changes. Likewise, when a DV change causes a process upset, the CV responds, and a new CV value is measured and becomes the controller's input. The error is now nonzero and a new MV value is calculated, the CV responds, and a new CV value is measured. This sequence continues until error approaches zero, the CV returns to its SP, and there are no further MV changes.

When an upset occurs, feedback controller corrective action begins only after the CV's value starts to deviate from its SP. The performance of feedback control can be improved by combining it with feedforward control,[9] a model-based control algorithm that provides corrective actions, e.g., new MV values, as soon as a DV change is detected, and before the CV begins to deviate from its SP. Because the DV is the feedforward controller's input, feedforward control is possible only for measured DVs or DVs that can be calculated or inferred from measured PVs. The basis for the feedforward controller's calculations are (linear) models of a CV's dynamic response to DV and MV changes; the models are developed using plant-test data. Both the feedback and feedforward corrections contribute to the MV corrections that bring the CV back to SP.

An example feedback control loop with feedforward control is shown conceptually in Figure 1.2; the relationships among a CV, MV, and DV are illustrated. Feedforward control essentials are presented in Chapter 3.

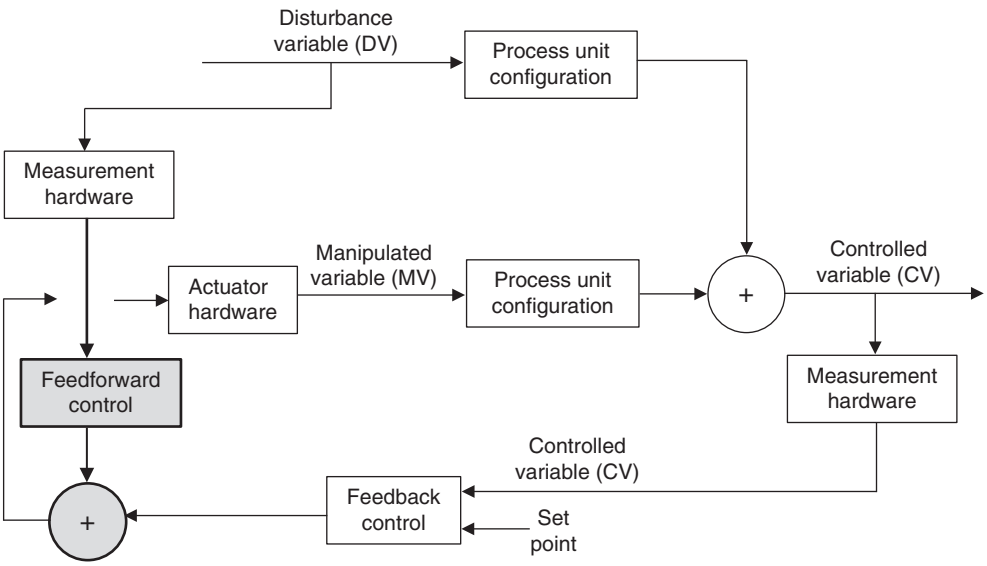


Figure 1.2 Feedback Control Loop with Feedforward Control

9 Feedforward control essentials are discussed in detail in Chapter 3.

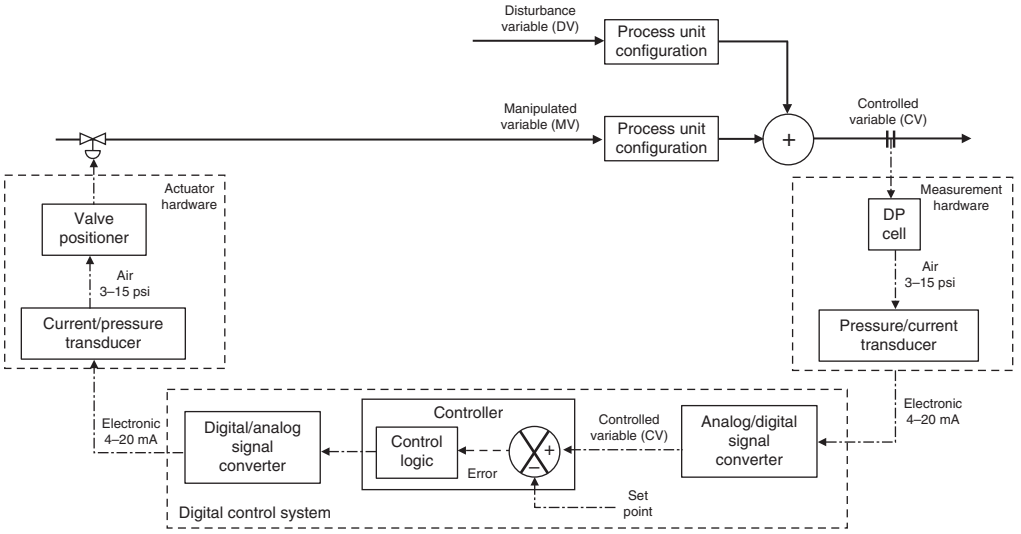


Figure 1.3 Control Loop Hardware

1.4.2 Control Loop Hardware

Several hardware components are essential to make any control loop functional. Figure 1.3 illustrates the hardware components and transmission path for a flow measurement and flow controller:

The measurement field hardware first generates a 3–15 psi air pressure signal, a pressure/current transducer converts it to an electronic 4–20 mA signal, and the 4–20 mA signal is transmitted to the control house where an analog/digital (A/D) signal converter generates the controller's digital input signal. The controller's calculated output is converted to a 4–20 mA signal by the digital/analog (D/A) signal converter and is transmitted to the actuator field hardware. A current/pressure transducer converts the 4–20 mA signal to a 3–15 psi air pressure signal that establishes the valve position. In most cases, the hardware components function transparently and are normally not considered during control strategy design.

However, hardware components can fail, and the failure risk of some components is higher than others. For example, many online analyzer measurements and externally multiplexed digital temperature indicators (DTIs) are more prone to failure than high-level *A*/*I* analog inputs (4–20 mA). The control strategy's designer must be aware of any potential hardware failures, especially those that can have detrimental process or control consequences. At a minimum, the designer must understand a failure's impact on the control strategy, and how the control strategy's response will impact the process. In many cases, user-provided logic/code is included in the control strategy's design to detect failures and take steps to prevent a process upset. The conventional advanced control applications presented in this book include methods to manage controller input and output failures using input/output signal validation methods.

1.4.3 Control Loop Examples

Three single control loop examples at light-ends tower T-85's overhead are illustrated in Figure 1.4. Overhead pressure PC77 PV is maintained constant at its desired target (PC77 SP) by manipulating air-fin condenser E-88's duty via its air louvre positioning. The flow of LVN product to storage FC86 PV is maintained constant at its desired target (FC86 SP) by flow controller FC86, and reflux

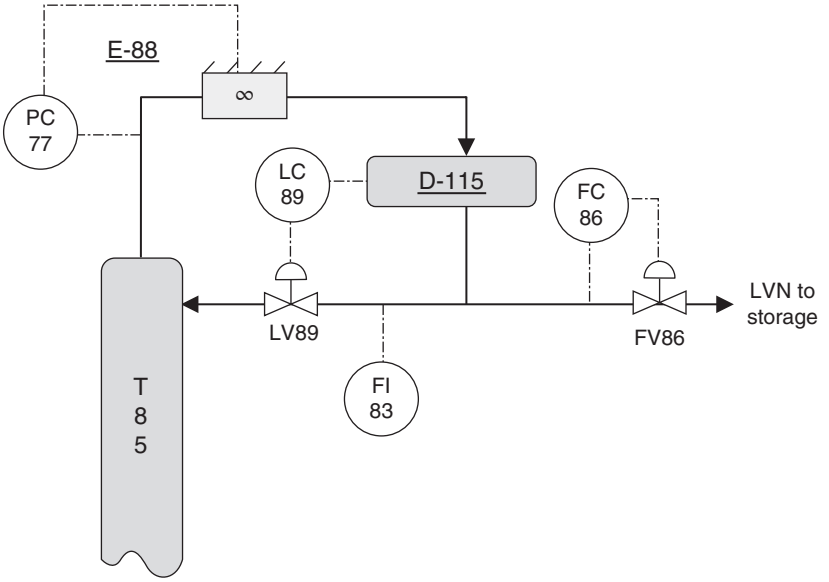


Figure 1.4 Control Loop Examples I – Single Loops

drum D-115's level LC89 PV is maintained constant by manipulating T-85's reflux return flow via valve LV89.

Additional control loops at the bottom segment of T-85 are shown in Figure 1.5. T-85 feed rate FC87 PV is maintained at its desired target (FC87 SP) by flow controller FC87. Tray temperature

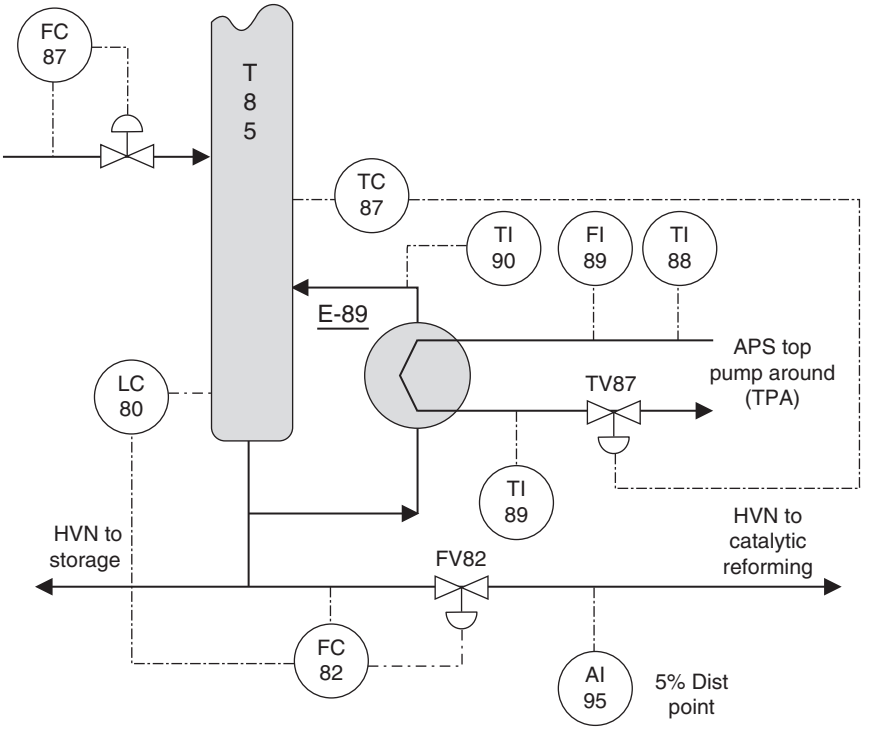


Figure 1.5 Control Loop Examples II – Single Loops and a Simple Cascade

TC87 PV is maintained constant at its desired target (TC87 SP) by manipulating reboiler E-89's duty via adjustments to valve TV87 to change CDU top pumparound (TPA) flow. Also, T-85's bottoms level LC89 PV is maintained constant by manipulating flow of bottoms heavy virgin naphtha (HVN) to catalytic reforming via adjustments to flow controller FC82's SP.

The LC80 → FC82 control cascade is an example of a simple cascade, e.g., one primary controller (LC80) sending its output to the SP of one secondary controller (FC82 SP). Complex cascades, with multiple primaries or multiple secondaries, are common in the design of conventional control application strategies. The structure of simple and complex cascades and cascade advantages and disadvantages are discussed in Chapter 2, Section 2.2.

1.4.4 Controller Modes

A controller's mode – Manual, Auto, or Cascade, determines whether the controller executes its control algorithm, who can change its SP and who can change its output. The following table summarizes a controller's functionality for each mode:

	Manual	Auto	Cascade
Control algo executed	No	Yes	Yes
Controller output or valve position (ultimate secondary) changed by	Operator	Control algo calculation	Control algo calculation
Controller SP changed by	N/A	Operator	Controller's primary

Therefore, in Figure 1.4, if controllers PC77, LC80, and FC86 are in Auto mode, all three control loops are commissioned, and their SPs can only be changed by operators. If an operator wants to change valve position FV86, FC86 must be transferred from Auto to Manual; however, flow control is no longer operational. In Figure 1.5, for the LC80 → FC82 cascade to be commissioned, LC80 must be in Auto and FC82 must be in Cascade. To change FC82 SP, an operator must transfer FC82 from Cascade to Auto. LC80 can no longer change FC82 SP, and level control is decommissioned. To change valve position FV82, FC82 must be transferred from Auto or Cascade to Manual mode. Just like transferring FC82 to Auto, LC80 can no longer change FC82 SP, and level control is decommissioned.

1.4.5 Open-loop Stable and Unstable Control Loops

When a controller is transferred from Auto to Manual mode, control calculations and controller outputs stop; the controller is in *open loop*. Control loops are open-loop stable if, in Manual mode, the CV lines out at a new steady state after an input variable change or disturbance. Control loops are open-loop unstable if, in Manual mode, the CV does not line out but keeps increasing or decreasing. *Almost all* control loops are open-loop stable or *self-regulating*.

Figure 1.6 presents an example of a self-regulating temperature control loop. In Figure 1.6, reactor R-401 feed rate (FI477) is set by feed surge drum D-425 level controller LC475. Reactor feed preheat, via a fired heater and heat integration, occurs *downstream* of the surge drum. After the catalytic reactions in R-401, reactor effluent light components (gas) are separated, and remaining liquid is fed to the product stripper where the final products are separated. Product stripper bottoms (hot side) and reactor feed (cold side) are heat integrated in exchangers E-420 A/B and via three-way bypass valve TV487, E-420 A/B duty is manipulated by TC487 to control D-425 inlet temperature. If TC487 is transferred to Manual mode, and an upset occurs that changes either E-420

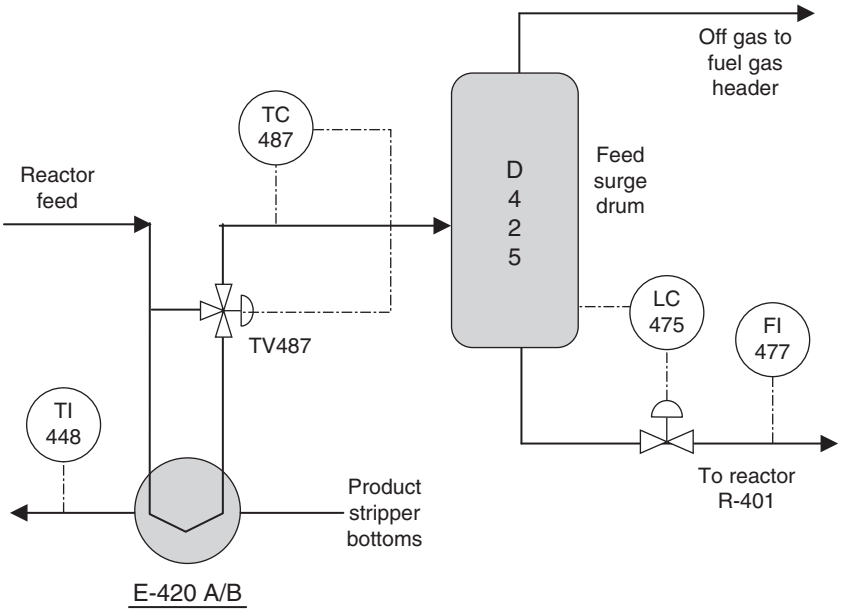


Figure 1.6 Open-loop Stable (Self-regulating) Temperature Control Loop

hot side (product stripper bottoms) or cold side (reactor feed) inlet temperature, D-425 inlet temperature will be upset but it will eventually line out at a new value as determined by an E-420 A/B heat balance. Because reactor feed preheat occurs *downstream* of D-425, R-401 inlet temperature is unaffected and can be maintained at its desired value; thus, a reactor conversion upset can be avoided.

The most common example of an open-loop unstable, non-self-regulating control loop is level control. In Figure 1.7, accumulator drum D-144 level LC122 PV is controlled by manipulating D-144 inlet flow via valve LV122. If at steady state, e.g., FI144 PV ≈ FC124 PV, LC122 is transferred to Manual mode, and an upset in inlet flow FI144 occurs, LC122 PV will either increase

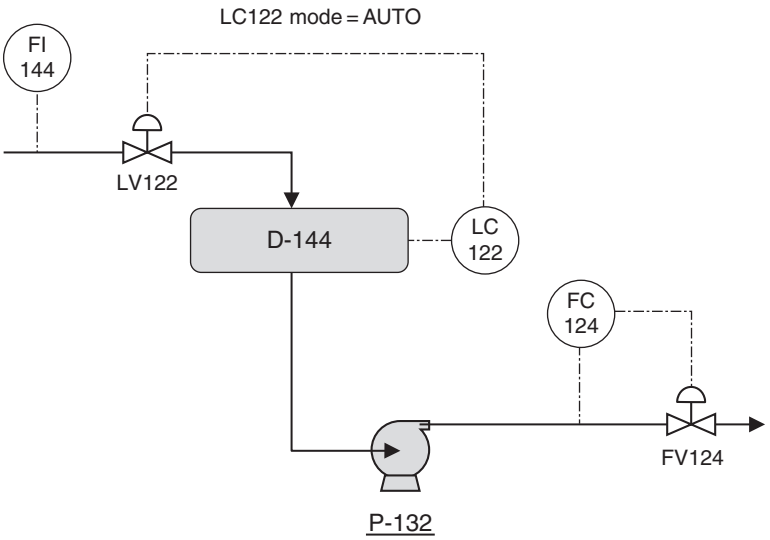


Figure 1.7 Open-loop Unstable (Non-self-regulating) Level Control Loop

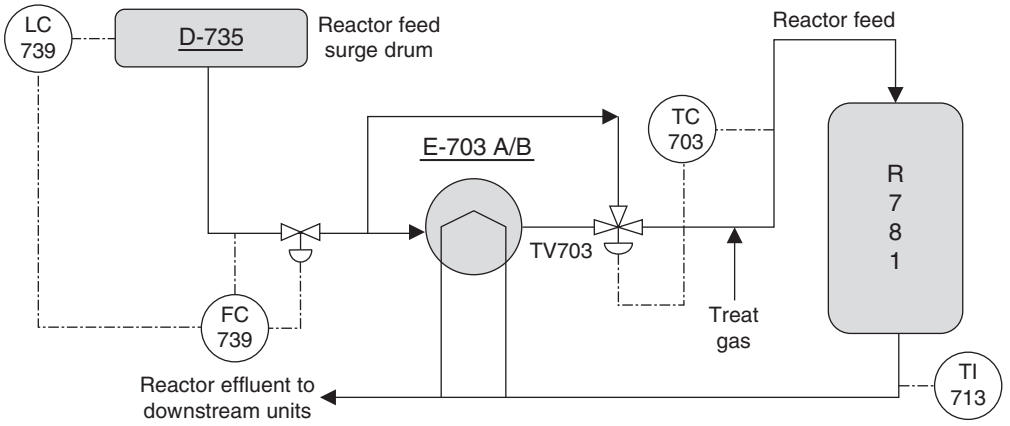


Figure 1.8 Open-loop Unstable (Non-self-regulating) Temperature Control Loop

(if inlet flow increases) or decrease (if inlet flow decreases) until D-144 inventory either overflows or empties. Unless D-144 outlet flow FC124 is changed, there is no *inherent* correction that can force the material balance to close at a new steady state.

If the feed to a reactor with exothermic reactions is preheated by hot reactor effluent, depending upon how reactor feed temperature is controlled, an open-loop unstable (non-self-regulating) temperature loop may exist. Such a process configuration is shown in Figure 1.8. R-781 is a packed multibed catalytic reactor in which exothermic reactions, in the presence of a hydrogen-rich treat gas to promote the desired reactions, produce the desired products. Although heat is generated and reactor bed temperatures rise as the reactions proceed along the catalyst beds, the reactor feed must be preheated to a high enough temperature to initiate the desired reactions. Inlet temperature controller TC703 maintains feed inlet temperature at a SP (TC703 SP) by manipulating E-703A/B duty that maintains the desired conversion. If TC703 is transferred to Manual mode, and a feed composition upset occurs that slightly increases the rate of reaction, because the reactions are exothermic, reactor bed temperatures rise. Thus, reactor effluent temperature increases, and after exchange with the hotter reactor effluent in E-703 A/B, feed temperature increases. This further promotes the exothermic reactions, rates of reaction increase further, and reactor bed temperatures continue to rise. This cycle of rising temperatures will continue, and since there is no *inherent* correction that can force the temperature to line-out at a new steady state, a dangerous temperature runaway can occur.

This temperature runaway can be avoided if the reactor effluent's measured temperature (TI713) is controlled and maintained at target. One possibility is to use a portion of the treat gas as interbed quench to control TI713 PV. However, additional hardware costs and reactor bed information is needed in order to assess whether a treat gas quench solution is feasible and/or justified.

1.5 Process Dynamics and Dynamic Responses

Control strategy design and troubleshooting plant upsets both require knowledge of process dynamic responses in the affected process units. In almost all cases, it is impractical to develop detailed, non-linear, first-principle dynamic simulations to study these dependencies. Instead, control engineers use process analysis, in many cases involving data collected during plant tests, to complete these tasks. The *process response curve* is an empirical tool that graphically shows

plant test results of a PV's (dependent variable) response to a change in another PV (independent variable). It provides visual insight into the dynamic relationship among PVs.

In the discussion of the feedback control loop (Section 1.4), there were many references to a CV's dynamic response to changes in an MV and a DV. The characteristics of the CV—MV dynamic response are central to a control loop's performance, and the suitability of a CV—MV pairing. The characteristics of CV—DV dynamic responses can provide helpful input to a control strategy's design, for example the need for feedforward control and other more advanced (and more complex) components. These CV—MV dynamic characteristics are now explored with the help of the process response curve.

1.5.1 Representation of Process Dynamics: Process Response Curves

Two basic characteristics that describe a dynamic response are *sensitivity* and *response time* or *settling time*. The *process gain* or *steady-state gain* (K_P) provides a measure of sensitivity, and the TTSS provides a measure of response time and settling time. The definition of K_P and the meaning of the TTSS are illustrated in Figure 1.9, which presents a typical dependent variable (e.g., a CV) dynamic response to a step change in an independent variable (e.g., an MV) of strength ΔX at time $t = 0$. This is an example of a process reaction curve. The dependent variable's response lines out at a new steady state, Y_{SS2}, at TTSS units of time after the initial change (at $t = 0$) in the independent variable. K_P, in EUs, is the ultimate change in the dependent variable, $\Delta Y = Y_{SS2} - Y_{SS1}$, divided by the change in the independent variable, $\Delta X = X_{SS2} - X_{SS1}$. The TTSS is the time it takes for the process response curve to essentially line out and remain at a new steady state value. *Note*: The estimation of K_P and TTSS requires the dependent variable to approach and remain at a new steady state value after the independent variable change.

In Section 1.5.2 we will discuss, illustrate, and quantify how process design and current operating conditions determine response dynamics. Just to provide an example, refer to feed surge drum D-425's process configuration in Figure 1.6. With inlet temperature controller TC487 manipulating

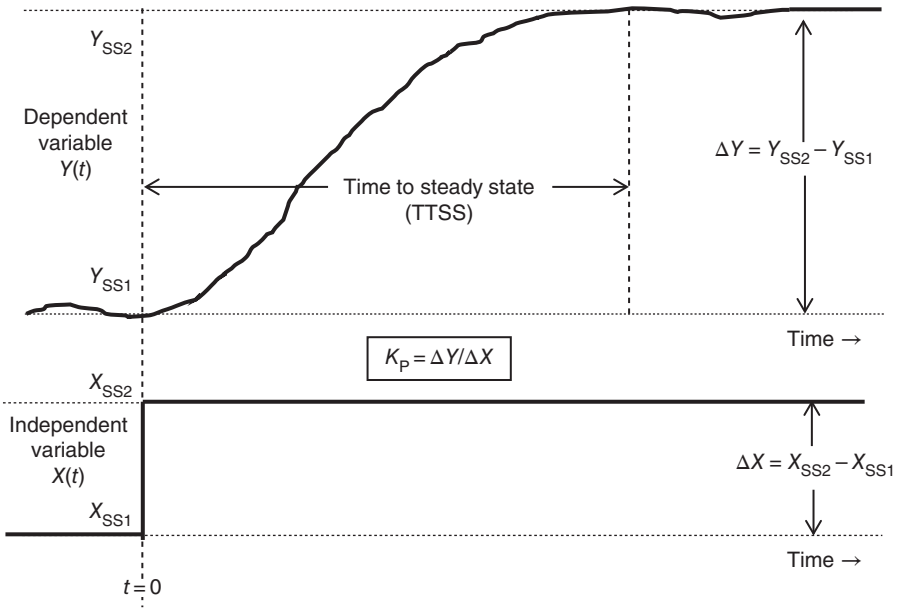


Figure 1.9 Dynamic Response Characteristics – Steady State Gain (K_P) and Time-to-steady-state (TTSS)

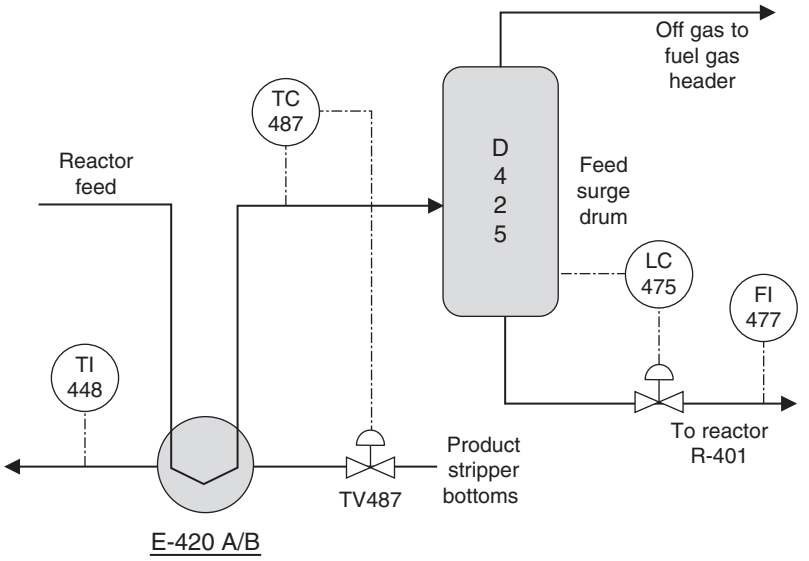


Figure 1.10 D425 Inlet Temperature Response Slowed by Exchanger Dynamics

three-way bypass valve TV487, the inlet temperature TC477 PV will respond to a TC487 SP change very quickly (small TTSS) because the dynamics of mixing are fast. With an alternative control valve type (less cost) and placement (hot side fluid flow to a product tank) shown in Figure 1.10, inlet temperature TC487 PV will respond to a TC487 SP change slower (larger TTSS) than the configuration in Figure 1.6 because E-420 A/B's slower (than mixing) dynamics are incorporated into TC487 PV's response.

The process response curve in Figure 1.9 does not indicate the presence of deadtime – the dependent variable's response starts just after the independent variable changes at $t = 0$. Depending upon its magnitude, CV–MV deadtime can make PID feedback control difficult and provide poor control performance. In many cases, long deadtimes are best managed using a single or multiple input – single output (e.g., multiple CVs and one MV) model-based control approach. To further our analysis of CV–MV response dynamics, Figure 1.11 shows a comparison of two dependent variable process response curves for changes in two different independent variables.

In Figure 1.11, unlike the process response curve for the change in the first independent variable (X_1), the process response curve for the change in the second independent variable (X_2) contains deadtime (θ_2). $TTSS_2$ is longer than $TTSS_1$ for two reasons: the presence of the deadtime and the slower, more ramp-like approach to its new steady state value after the deadtime expires.

Remember, CVs are dependent variables and MVs are independent variables, so now consider the response curves in Figure 1.11 as CV responses to step changes in two different potential MVs, starting from the same initial steady state value ($Y_{1,SS1} = Y_{2,SS1}$). Qualitative but useful guidelines to help judge whether an MV is suitable for the control of a specific CV are (i) the magnitudes of K_P and TTSS – the larger the K_P value and the smaller the TTSS value, the more suitable the MV choice and (ii) the existence and magnitude of deadtime in the CV response – the larger the deadtime, the more troublesome closed-loop control. In Figure 1.11, the CV–MV_1 pairing has a shorter TTSS and no deadtime, both favoring it over MV_2 as a potential MV. Notice that the TTSS includes the deadtime duration, but the presence of deadtime is in itself an important factor above and beyond the longer TTSS. As for the steady state gains, K_P, they appear to have comparable magnitudes => with $\Delta Y_1 > \Delta Y_2$ and $\Delta X_1 > \Delta X_2$, the ratios $\Delta Y_1/\Delta X_1$ and $\Delta Y_2/\Delta X_2$ will have similar magnitudes.

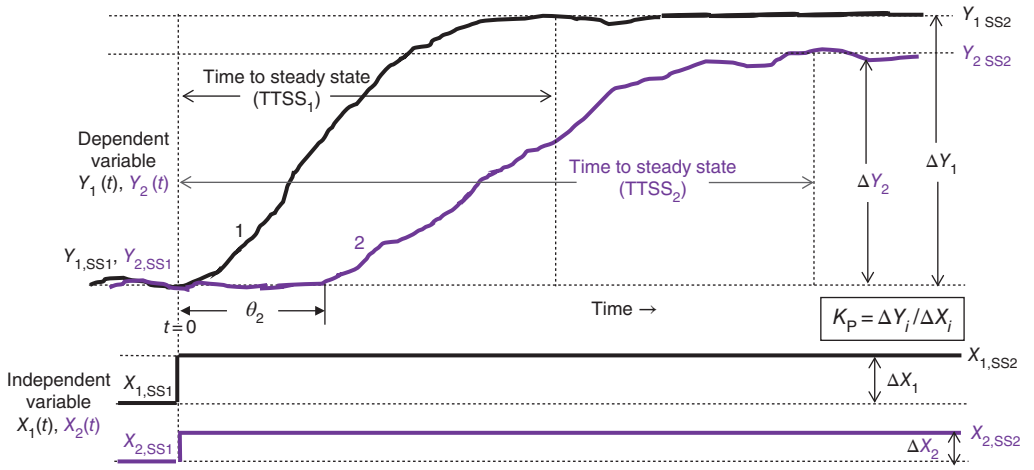


Figure 1.11 Dependent Variable Responses to Changes in Two Independent Variables

Another factor to consider in the CV–MV pairing issue is interactions among CVs and MVs. In many cases, a specific MV will have impact on more than one CV and multiple MVs will have an impact on a single CV. As discussed above, if multiple MVs have an impact on a single CV, process analysis and plant testing facilitate making a reasonable MV choice. Assume CV_1 is paired with MV_1 and CV_2 is paired with MV_2, but MV_1 also impacts CV_2 and MV_2 also impacts CV_1. Then CV_2's control loop will be impacted anytime MV_1 takes corrective action to control CV_1. Likewise, CV_1's control loop will be impacted anytime MV_2 takes corrective action to control CV_2. The two control loops interact, and the performance of both loops will degrade. As we will observe when we present and discuss industrial applications, multivariable interactions, and dynamic responses with long deadtimes are often better managed by a multivariable constraint control (MVCC) method such as DMC.

Now consider the CV response curves in Figure 1.11 as responses to step changes in two different DVs (remember DVs are also independent variables), instead of two different MVs. Plant test data can provide a qualitative guideline, which may be helpful, to assess whether feedforward control is needed for a specific DV. If the CV–DV response curve has small K_P and a long TTSS, the impact of the DV's change on the CV's response would appear to be minor (small K_P) and affect the CV over an extended time period (long TTSS). If so, feedback control alone may be able to effectively manage the upset, and feedforward control may not be necessary. *Note*: The detailed discussion of feedforward control in Chapter 3 will indicate that in addition to the CV–DV response dynamics, the CV–MV response dynamics impact the feedforward controller's design.

Operators, who in their daily tasks constantly observe and interact with the process, normally have a good "feel" for the effect changes in their units' PVs have on process operations. They can thus provide valuable input on dynamic cause and effect issues, e.g., response times and PV sensitivities, and important upsets and their consequences. In almost all cases, discussing these issues with operators is helpful.

1.5.2 First-Order Linear Continuous Model

Up to this point we have been using process response curves to describe and understand the dynamic relationships among CVs, MVs, and DVs. Keep in mind the industrial control applications presented in this book focus on the impact process configuration, operating objectives, and control

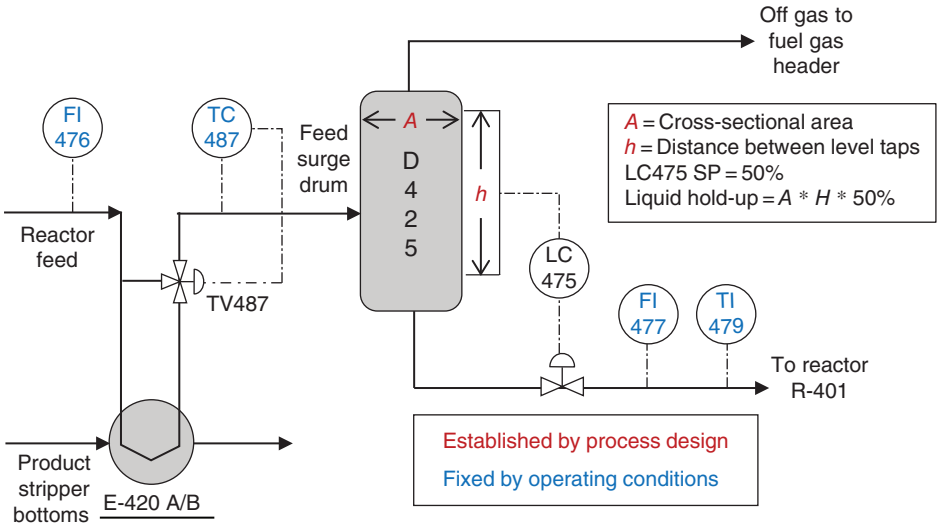


Figure 1.12 D-425 Process Design and Operating Conditions That Define Process Dynamics

objectives have on a control strategy's design. Control strategy design for the most part does *not* depend on or need dynamic simulations to provide realistic, mathematically based results. Instead, a less fundamental but still meaningful formulation of process dynamics is sufficient for control strategy design. We now relate this formulation to the underlying concepts of process dynamics provided visually by the process response curve.

We will use feed surge drum D-425, first introduced in the discussion of Figure 1.6, as the basis for developing a simple (linear) dynamic model; Figure 1.12 shows D-425's process configuration. There is additional information included in Figure 1.12, but absent from Figure 1.6. The cross-sectional area (A), the distance between level taps (h), and the level controller's set point (LC475 SP) are needed to calculate the liquid hold-up in D-425. Inlet (FI476) and outlet (FI477) flows and outlet temperature (TI479) are needed (along with liquid inventory) for a D-425 heat balance. Note that the cross-sectional area and distance between level taps are established by process design, but the values of inlet and outlet flows, inlet and outlet temperatures, and the level controller's SP are set by process operating conditions.

We now derive an expression that mathematically describes the response of outlet temperature TI479 PV ($\equiv T_{OUT}$) to a change in the inlet temperature set point ΔTC487 SP ($\equiv \Delta T_{IN}$). We assume D-425 is well mixed and is initially at steady state, e.g., there is no material (FI476 PV = FI477 PV ≡ F) or energy ($T_{IN} = T_{OUT}$) accumulation in D-425. With liquid holdup ≡ $V = A * h *$ (LC475 SP), a D-425 unsteady state heat balance yields:

$$\tau * \frac{dT_{OUT}}{dt} = [T_{IN}(t)] - [T_{OUT}(t)] \tag{1.1}$$

$$\text{where: } \tau = \frac{V}{F} = \frac{\text{Holdup}}{\text{Throughput}} \tag{1.2}$$

Equation (1.1) is a linear, first-order differential equation; the time domain solution, $T_{OUT}(t)$, to a step change in T_{IN} of strength ΔT_{IN} is given by:

$$T_{OUT}(t) = T_{OUT}(0) + \Delta T_{IN} * \left(1 - \exp\left[-t/\tau\right]\right) \tag{1.3}$$

where $T_{OUT}(0)$ is the value of T_{OUT} when T_{IN} changes. In Equation (1.3), τ is the *first order time constant*, defined in Equation (1.2), which determines how fast the outlet temperature approaches its

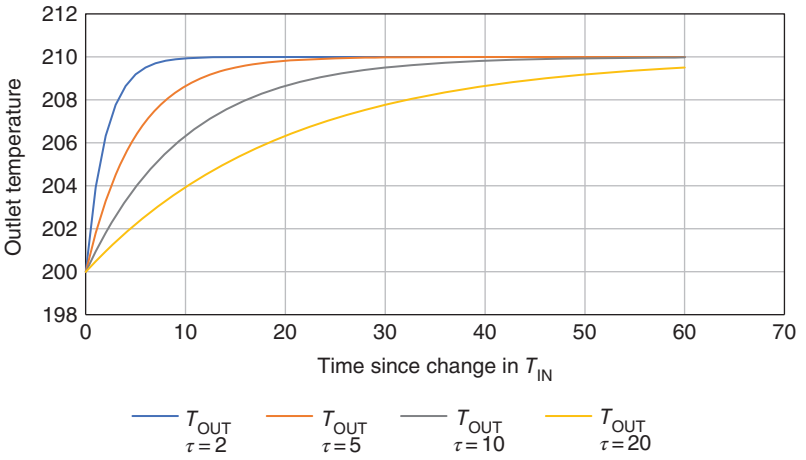


Figure 1.13 D-425 Outlet Temperature Responses for Four Values of τ (Minutes)

new steady state value, and an indication of its TTSS. From its definition in Equation (1.2), the outlet temperature response's time constant (τ), and hence its TTSS, (i) increases with increasing holdup or decreasing throughput and (ii) decreases with decreasing holdup or increasing throughput. Temperature responses for four values of τ (2, 5, 10, and 20 minutes) are presented in Figure 1.13. These responses were obtained for an inlet temperature TC487 SP change at $t = 0$ of 10 °C, from 200 → 210 °C. These results clearly indicate response times get slower, and TTSS gets longer, as τ increases. These results also indicate that each response's TTSS can be estimated as approximately equal to five time constants.

Note that in the above example, there is no chemical reaction or external heat source or sink to change T_{OUT} other than to respond to the T_{IN} change. Therefore, $K_P \equiv 1$, and in fact a 10° increase in T_{IN} results in a 10° increase in T_{OUT}.

The linear, first-order differential equation in Equation (1.1) can be made more general by writing it for any dependent variable, Y, for the change in any independent variable, ΔX, and including steady state gain, K_P, and deadtime θ:

$$\tau * \frac{dY(t)}{dt} = K_P * u(t-\theta) * \Delta X(t) - Y(t) \tag{1.4a}$$

where the unit step change $u\ (t–\theta) = 0$ for $t < \theta$ and $u(t–\theta) = 1$ for $t \geq \theta$. In effect, this shifts the start of the independent variable's step change, $\Delta X(t)$, θ units later in time. Given the properties of the unit step $u(t–\theta)$, the product $u(t–\theta) * \Delta X(t) \equiv 0$ for $t < \theta$ and $u(t–\theta) * \Delta X(t) \equiv \Delta X(t)$ for $t \geq \theta$; this is indicated by writing this product as $\Delta X(t–\theta)$. With this change in notation, Equation (1.4a) becomes:

$$\tau * \frac{dY(t)}{dt} = K_P * \Delta X(t-\theta) - Y(t) \tag{1.4b}$$

The time domain solution of Equation (1.4b) is:

$$Y(t) = Y_{SS1} + K_P * \Delta X(t-\theta) * \left(1 - \exp\left[-\frac{(t-\theta)}{\tau}\right]\right) \tag{1.5a}$$

where $\Delta X \equiv X_{SS2}–X_{SS1}$, and X_{SS1} and Y_{SS1} are the steady state values of X and Y for $t < \theta$. Rearranging Equation (1.5a) to solve for $\Delta Y \equiv Y(t) – Y_{SS1}$ yields:

$$\Delta Y(t) = K_P * \Delta X(t-\theta) * \left(1 - \exp\left[-\frac{(t-\theta)}{\tau}\right]\right) \tag{1.5b}$$

Remember $\Delta X\,(t–\theta) = 0$ for $t < \theta$ and $\Delta X\,(t–\theta) = \Delta X(t)$ for $t \geq \theta$. In effect, the presence of deadtime shifts the dependent variable's response θ units later in time, e.g., θ units of time *after* the independent variable's change at $t = 0$. Applied to the responses in Figure 1.13, all responses would shift to the right, θ units later in time, but the steady state gain and shapes of the curves would be unchanged.

An estimate for the first-order time constant, τ, can be obtained from Equation (1.5b). In Equation (1.5b), when $t > \theta$ and $t = \theta + \tau$:

$$\text{If } t > \theta\text{: } \Delta X\,(t-\theta) = \Delta X(t) = \Delta X = X_{SS2} - X_{SS1}$$

$$\text{If } t = \theta + \tau\text{: } \left(1 - \text{Exp}\left[-\frac{(t-\theta)}{\tau}\right]\right) = (1 - \exp[-1]) = 0.63$$

And, remember, $K_P * \Delta X \equiv \Delta Y = Y_{SS2} - Y_{SS1}$, and Equation (1.5b) becomes:

$$\Delta Y(t = \theta + \tau) = K_P * \Delta X * 0.63$$

$$\Delta Y(t = \theta + \tau) = 0.63 * \Delta Y = 0.63 * (Y_{SS2} - Y_{SS1}) \tag{1.6}$$

Thus, for a first-order linear dynamic system, the time *after the deadtime's duration,* or if $\theta = 0$, the time at which the dependent variable reaches 63% of its ultimate change (ΔY) is an estimate of the first-order time constant, τ.

True Deadtime

If, for some reason, the outlet temperature measurement sensor can only be placed well downstream of D-425's outlet, T_{OUT} would not start to change immediately after the change in T_{IN}. Instead, the start of the T_{OUT} response would be delayed until the first elements of D-425's effluent stream with an elevated temperature reaches the sensor. This is an example of a "*true*" *deadtime*, whose magnitude can be estimated from pipe geometry and D-425's outlet flow:

$$\theta = (\text{Pipe Cross} - \text{Sectional Area} * \text{Distance to Sensor})/\text{Outlet Flow}$$

Remember that with dead time present, all outlet temperature response curves appearing in Figure 1.13 would be shifted θ time units to the right, but the *steady state gain and shapes of the curves would not change*.

Assume there is no heat loss, reaction, or mixing (liquid is in plug flow) along the path to the temperature sensor. Starting with Equation (1.5b), once again evaluate $\Delta Y(t)$ when $t > \theta$ and $t = \theta + \tau$. The term (1–exp [− $(t–\theta)/\tau$]) becomes (1–exp [−1]), which again is equal to 0.63. When $t > \theta$, $\Delta X\,(t–\theta) = \Delta X(t)$, the magnitude of the independent variable's step size, $\Delta X = X_{SS2}–X_{SS1}$, and the dependent variable's new steady state value, ΔY, is equal to $K_P * \Delta X$. Equation (1.5b) becomes:

$$\Delta Y(t = \theta + \tau) = 0.63 * \Delta Y = 0.63 * (Y_{SS2} - Y_{SS1}) \tag{1.7}$$

Thus, consistent with Equation (1.6), Equation(1.7) indicates the time *after the deadtime's duration* at which the dependent variable reaches 63% of its ultimate change ($\Delta Y = Y_{SS2} - Y_{SS1}$) is an estimate of the first-order time constant, τ. This is illustrated in Figure 1.14.

Apparent Deadtime

True deadtime is real and is present because of the process configuration. *Apparent* deadtime will arise when a real process response curve is approximated by a first-order linear response. In Figure 1.15, the dependent variable's actual process response initially decreases, and then eventually turns and begins to increase; it continues to rise, and asymptotically approach its ultimate

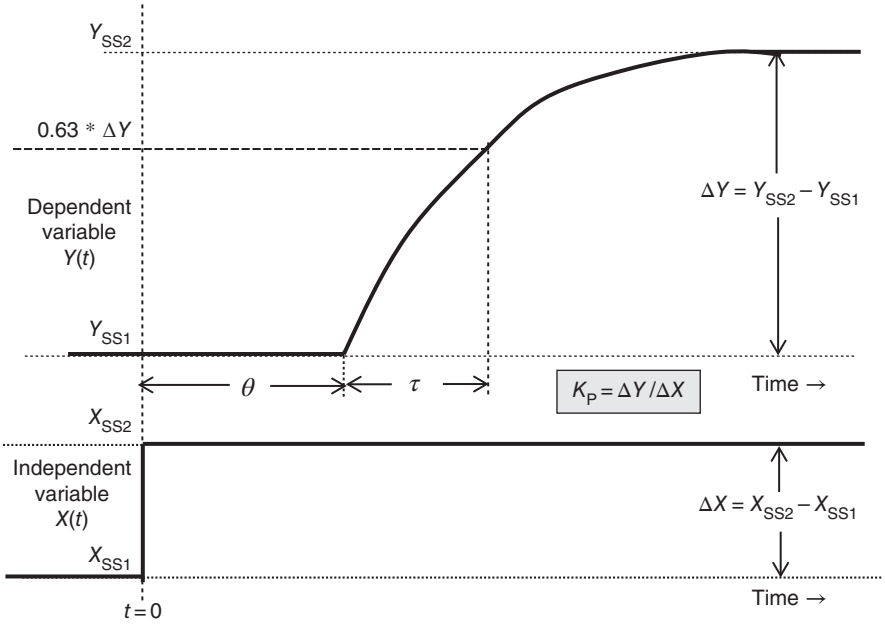


Figure 1.14 Estimating First-Order Time Constant (τ) for Response with Deadtime

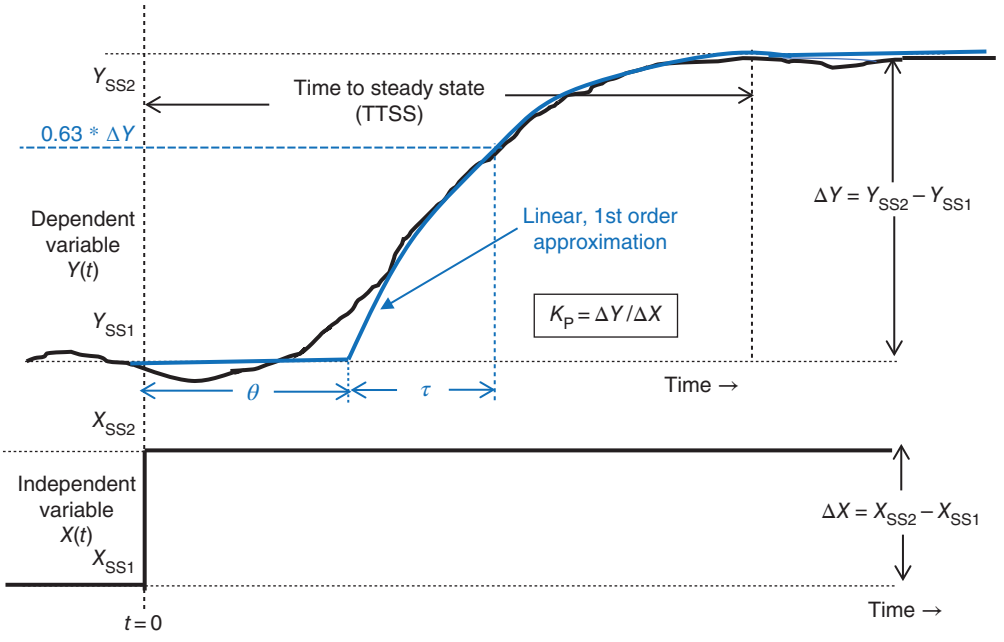


Figure 1.15 Estimating Apparent Deadtime

steady state value. When a dependent variable's response to a step change moves initially in one direction but ultimately reverses and moves in the opposite direction before it approaches its new steady state, the response is called an *inverse response* (Ref. [1]). If the dependent variable is a CV, its MV control adjustments will initially be in the wrong direction, make the situation worse, and negatively impact control performance. Of course, choosing to approximate this inverse response

with a first-order approximation will *not* improve closed loop control performance; however, the first-order approximation's model parameters (K_P, τ, and θ) can be useful for PID tuning[10] and necessary for single input – single output model-based control.[11]

As shown in Figure 1.15, the first-order approximation substitutes an interval of deadtime for the initial inverse response, and as discussed in Section 1.5.2.0 (True Deadtime), the first-order time constant τ is estimated using the 63% guideline. The first-order approximation is not particularly good for $0 \le t \le \theta$, but for $t > \theta$, it is reasonable. Also, the dependent variable's ultimate steady state value is not affected by the model errors at $t \le \theta$; thus, the first-order approximation can be made to asymptotically approach the correct, new steady state value, and the approximation's K_P estimate is reasonably accurate.

Apparent deadtime will often be observed in the plant because process operations in the real world are complex. The overall process responses observed in the plant are complex combinations of individual unit, measurement, and actuator responses. Also, most processes are nonlinear, distributed (PVs are functions of time *and* distance), and their dynamic responses are functions of operating conditions. Therefore, in many cases, the observed responses are slow to start, and compared to the initial fast rise in the dependent variable's response in Figure 1.14, responses observed in the plant will have more rounded initial rises. The initial portion of the response will often look like apparent deadtime. As already noted, and as illustrated in Figure 1.15, a linear model used to approximate a real-world process response will almost always substitute an interval of apparent deadtime for the initial time interval where there is little observed changes in the dependent variable.

1.5.3 Transfer Function Representation

A *transfer function* is an empirical model that relates an output variable – a dependent variable, to an input variable – an independent variable, not in the time-domain but in the Laplace domain, or s-domain. As included in all undergraduate process control textbooks (Ref. [2, 3]), Laplace transforms convert linear ordinary differential equations (with constant coefficients) to algebraic equations. Examples would be unsteady state material and energy balances, with a structure like Equations (1.4a) and (1.4b). Taking the Laplace transform of both sides of Equation (1.4b) yields:

$$\tau * s * Y(s) - Y(t < \theta) = K_P * \exp(-\theta s) * X(s) - Y(s) \tag{1.8}$$

Since $Y(t < \theta) \equiv 0$, it can be dropped from Equation (1.8); re-arranging and solving for $Y(s)$:

$$Y(s) = \left[\frac{K_P * \exp(-\theta s)}{\tau s + 1} \right] X(s) \tag{1.9}$$

The relationship between dependent ($Y(s)$) and independent ($X(s)$) variable Laplace transforms is defined as the *transfer function*, $G(s)$:

$$G(s) \equiv \frac{Y(s)}{X(s)} = \left[\frac{K_P * \exp(-\theta s)}{\tau s + 1} \right] \tag{1.10}$$

The process response curve in Figure 1.14 is an empirical, graphical representation of the solution (the dependent variable's response) of a linear, ordinary first order differential equation (such as Equation (1.4b) to a step change in an independent variable. Equation (1.5b) is an equivalent analytical time domain representation of the solution of the linear first order ordinary differential equation in Equation (1.4b), while Equation (1.10) is an equivalent representation of this solution in the s-domain. Therefore, all three are equivalent representations of the dynamic response, e.g., the solution to Equation (1.4b).

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10 See Section 2.4.7 for a discussion of PID tuning.
11 See Section 2.7.

There are times when a higher order linear approximation of plant dynamics is a better representation of those dynamics than the first order transfer function in Equation (1.10). A second order transfer function includes additional dynamic parameters that enable it to better represent several types of process responses, for example, those that include inverse response and/or overshoot. Here is a common form of a second-order transfer function:

$$G(s) = \frac{Y(s)}{X(s)} = \frac{K_P * (1 + \tau_3 s) * \exp(-\theta s)}{(1 + \tau_1 s) * (1 + \tau_2 s)} \tag{1.11}$$

The second order transfer function's model parameters are the time constants τ_1, τ_2, and τ_3, the deadtime θ, and the steady state gain, K_P. The τ_3 term in the numerator tends to speed the response, while the τ_1 and τ_2 terms in the denominator tend to slow the response. If $\tau_3 < 0$, the second order transfer function in Equation (1.11) can provide an initial inverse response for an independent variable's response; if $\tau_3 > \tau_1 \geq \tau_2$, it can provide an overshoot in an independent variable's response (Ref. [4]).

If $\tau_2 = 0$ in Equation (1.11), the resulting transfer function is called a "lead-lag" and is the basis for understanding the feedforward control algorithm:

$$G(s) = \frac{Y(s)}{X(s)} = \frac{K_P * (1 + \tau_3 s) * \exp(-\theta s)}{(1 + \tau_1 s)} \tag{1.12}$$

The time constant τ_3 in the numerator is called the "lead" time constant, while the time constant in the denominator τ_1 is called the "lag" time constant. The effect these two time constants have on the response in Equation (1.12), and their significance in the feedforward control algorithm, will be explained and discussed in detail in Chapter 3.

1.5.4 First-order Linear Discrete Model

Up to now, we have represented process dynamics with continuous models in the time domain. Equation (1.4b) is a first order linear ordinary differential equation with time-domain solution given in Equations (1.5a) and (1.5b). The solution in the *s*-domain, equivalent to the time domain solution in Equations (1.5a) and (1.5b), is the first order transfer function in Equation (1.10). In addition, a second order transfer function, equivalent to the time domain solution of a second order linear ordinary differential equation, is given in Equation (1.11).

In today's DCS world, data is *sampled* at a fixed frequency and process information is known only at discrete times. This is illustrated in Figure 1.16:

The current value of the dependent variable at time step "*n*," Y_n, is considered constant between sample times; as shown in Figure 1.16, the dependent variable becomes a "staircase" function, as do independent variable inputs, e.g., MVs and DVs. Linear ordinary differential equations and *s*-domain transfer functions are not valid in the discrete world, and a different, a *discrete* representation of the process reaction curve is needed. The discrete analogue of differential equations and time domain solutions are recursive difference equations that contain dependent and independent variable values at specific time steps, e.g., Y_n, Y_{n-1}, and X_{n-1}. The discrete analogue of *s*-domain transfer functions are *z*-transforms. The difference equation representation of the process response curve is the "inverse *z*-transform" just as the time domain expression in Equation (1.5b) is the "inverse Laplace (*s*)-transform" of Equation (1.9).

The mathematical relationships among the sampled function (Y_n), the continuous function ($Y(t)$), the definition of the *z*-transform, the inverse *z*-transform, and the difference equation representation of the process response curve are complex (Ref. [5]) and are not relevant to real-world control strategy design. Therefore, only an important example is provided below, e.g., a first-order sampled data difference equation with deadtime, to introduce the recursive nature of

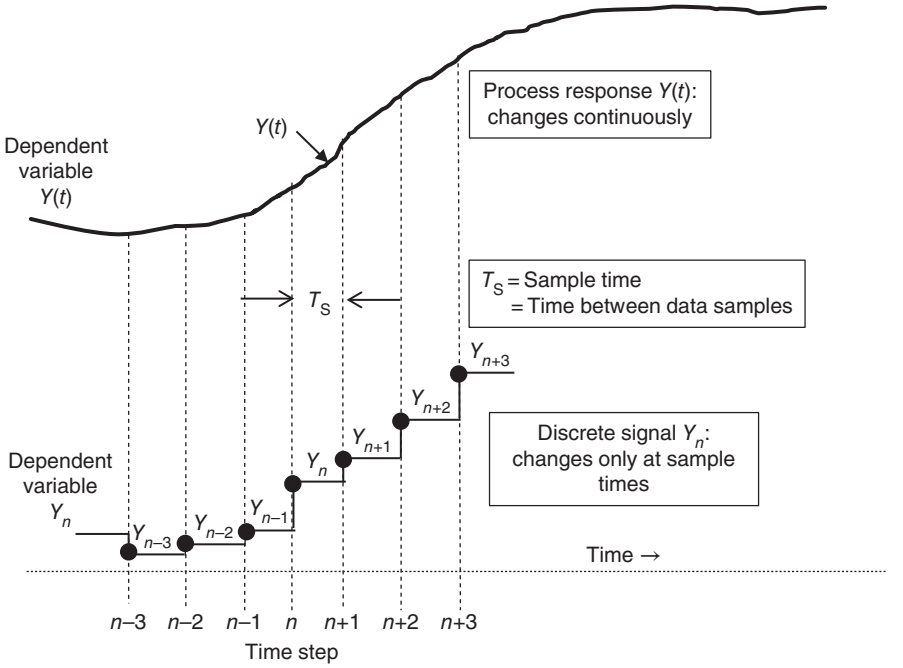


Figure 1.16 Continuous and Discrete Signals

the discrete representation of the process response curve. This will be helpful when the standard PID control algorithm (Section 2.4) and the standard linear exponential filter (Section 2.5), both recursive, difference equations, are presented and discussed.

$$Y_n = b * Y_{n-1} + (1-b) * K_P * X_{n-1-\Gamma} \tag{1.13a}$$

$$b = \exp(-T_S/\tau) \tag{1.13b}$$

$$(1-b) = 1 - \exp(-T_S/\tau) \tag{1.13c}$$

$$\Gamma = \theta/T_S \tag{1.13d}$$

Sample times (T_S) are normally chosen small enough that the ratio of deadtime to sample time, Γ, can be assumed to be an integer. For example, with $T_S = 4$ seconds and $\theta \approx 87$ seconds, then $\Gamma = 21.8$ which can be rounded up to ≈22. To interpret the practical meaning of Equation (1.13a), assume $T_S = 5$ seconds and $\theta = 60$ seconds; then, $\Gamma = 12$ and Equation (1.13a) becomes:

$$Y_n = b * Y_{n-1} + (1-b) * K_P * X_{n-13}$$

In words, the current value Y_n depends upon its value one-time step ago, Y_{n-1}, and to account for the deadtime of 12 time steps, the value of input X 13 time steps ago; this is the *last* value of X that has affected the value of Y_n at the current time step, n.

In Section 2.5, measurement noise and the need for signal filtering will be discussed. Almost all DCS' include a *linear exponential filter* as a standard filter algorithm – the most widely used filtering algorithm. It is very closely related to first order sampled data difference equation in Equation (1.13a). For no deadtime ($\Gamma = 0$), $K_P = 1.0$, and τ in Equation (1.13b) set equal to τ_f, the filter time constant, Equation (1.13a) becomes:

$$Y_n = b * Y_{n-1} + (1-b) * X_n \tag{1.14a}$$

$$b = \exp(-T_S/\tau_F) \tag{1.14b}$$

With Y signifying the filtered value, and X the raw measurement, Equation (1.14a) is the recursive difference equation for the linear exponential filter. *Important*: Unlike the standard first order sampled data difference equation in Equation (1.13a), to let the filter respond faster to a raw measurement change, the linear exponential filter in Equation (1.14a) uses the *current* raw measurement, X_n, not the previous measurement, X_{n-1}.

1.6 Plant Testing

Many conventional control application designs are presented and discussed throughout this book. The most important requirement for developing a good control strategy design is an in-depth understanding of the process – the unit configuration, operating objectives, economic drivers, control objectives, and available measurements and control hardware (e.g., valves). In addition, potential constraints, potential upsets, potential DVs, CVs, and MVs, and important interactions and sensitivities among MVs, CVs, and DVs must be identified and understood.

Plant tests establish how planned changes in independent variables dynamically affect one or more dependent variables. A plant test may be as simple as changing the SP of one independent variable (an MV or a DV) and collecting response variable data for one dependent variable (a CV). The dependent variable responses to changes in two independent variables shown in Figure 1.11 are examples. On the other hand, plant tests to develop a DMC model consist of a carefully planned sequence of changes in *multiple* independent variables (both MVs and DVs) and the collection of dynamic response data for *multiple* dependent variables (CVs).

1.6.1 Objectives

Plant tests are performed to increase the understanding of dynamic relationships among PVs in a specific process area. Data collected during a plant test, and the data analysis that follows are used to (i) identify important dynamic relationships among potential CVs, MVs, and DVs, (ii) identify and assess potential control interactions, (iii) evaluate steady state sensitivities, and (iv) provide model parameter values that are suitable for both model-based control and PID tuning. All such information can be used to resolve relevant technical issues, and then help weigh and evaluate competing control strategy designs. Although not directly relevant to process control, plant tests are routinely used to troubleshoot process problems and/or to identify the root-causes of process upsets. *Important*: It is essential to check data collection hardware and software *before* testing begins to ensure they are fully operational.

1.6.2 Work Process

Plant tests have an impact on plant operations and therefore it is essential that disruptions to process operations are minimized. Plant test objectives and procedures should always be discussed with console operators prior to the start of the test. Operator experience is invaluable when choosing the move (step) size to ensure the dependent variable changes are large enough to be measured and analyzed. Operators can also suggest a reasonable TTSS that ensures sufficient time is allotted for the dependent variable to line out at a new steady state value. The basis for the plant test, e.g., the assumed relationships among PVs, should be discussed with process technical personnel.

Noise and filters will be discussed in Chapter 2, Section 2.5. However, if noise on a dependent variable appears to be a problem, then filter the noisy signal *before* plant test data is collected. This ensures τ_f, the filter time constant, is incorporated into the dependent variable's dynamic response

and thus into the model's dynamic parameter values. The signal-to-noise ratio, a parameter that qualitatively estimates the strength of the base (process) signal versus the magnitude of the noise will also be discussed in Section 2.5. If a dependent variable is noisy, even if it is filtered, the independent variable's step change magnitude must be large enough to ensure the dependent variable's response is indicative of the base signal and not the noise.[12]

It is best to work at the operator's console with operators as the test proceeds. It is important that operators not make routine process changes that impact the same dependent variable(s) whose responses to planned independent variable step changes are being studied. In addition, tests should be scheduled at a time when operations are normal – no feed rate or feed composition changes expected, and process operations in the specific unit are initially at or close to steady state.

Lastly, control loops that have impact on the plant test's dependent variables should be identified. They must be on control (in Auto or Cascade mode), not wound-up, and tuned well. If not, these problems should be resolved, controllers retuned if necessary, and put on control before commencing the plant test. If these problems are not resolved, the dependent variable's response dynamics will not incorporate the normal impact of the underlying controls. Consideration should be given to monitor the underlying controller status and performance during the plant test. Also, it is beneficial to identify potential process disturbances that can affect the dependent variable responses. Their impact must be understood; if possible, take proper precautions, and decide what actions will be taken if they occur. It is usually a good idea to collect data for all PVs of possible interest in the plant test's focus area.

1.6.3 In Process Control

Look again at Figure 1.11 – it illustrates how dependent variable (CV) responses to step changes in two different independent variables (MVs) can provide useful guidelines for pairing a CV with an MV. By comparing the two responses, the larger the K_P value (better sensitivity) and the smaller the TTSS value (faster the approach to a new steady state), the more suitable the MV choice. Also, the larger the deadtime, the more troublesome closed-loop control will be. In Figure 1.11, the CV–MV_1 pairing has a shorter TTSS and no deadtime, both favoring it over MV_2 as a potential MV. *Remember*: The K_P and TTSS parameter estimates are dependent upon the CV approaching and lining out at a new steady state value after the MV step change.

Also, look again at Figure 1.14; we previously discussed a simple graphical procedure to estimate a first order linear model's dynamic parameters, K_P, τ, and θ. But the value of τ is based on only *one* data point, the time at which the dependent variable reaches 63% of its ultimate change. Parameter values estimated in this way are *minimally* suitable for model-based control but may be acceptable to use to estimate PID tuning constants. More effective graphical methods are available that use *two* points on the response curve to estimate τ and θ (Ref. [2, 3, 6, 7]). This two-point graphical method can provide parameter estimates that are acceptable for single or multiple input – single output model-based control, for tuning PIDs and for identifying feedforward controller lead time and lag time constants.

Commercially available model identification tools (Ref. [8]), which use statistical curve fitting techniques, are far more effective at obtaining accurate model parameter values compared to these graphical methods. A model is a mathematical representation of a process response curve. A model has a *structure*, e.g., linear, nonlinear, first or second order with or without deadtime, and a set

12 This depends on the signal-to-noise ratio, discussed in Section 2.5.

of dynamic parameters. Model identification tools find values for model parameters that provide the "best" fit of a specific model to the plant test data. In many cases, plant tests are run and model identification tools used to fit response data to a first-order (Equation (1.10)), or a second order (Equation (1.11)) linear model. Model identification tools use various statistical methods to calculate the best set of model parameters, but no matter the statistical methods used, they all calculate and use the error, or *residual* between predicted PV values ($Y_{i,\text{Pred}}$) and measured PV values (Y_i)–Residual$_i \equiv (Y_i - Y_{i,\text{Pred}})$. The curve-fitting technique calculates the best set of model parameter values, i.e., the set of model parameter values that minimizes the sum of squares of the residuals:

$$\text{Minimum} \mid_{\substack{\text{Dynamic} \\ \text{Parameters}}} = \sum_{i=0}^{N} (Y_i - Y_{i,\text{Pred}})^2 \tag{1.15}$$

Y_i = Dependent variable plant test data point
$Y_{i,\text{Pred}}$ = Model's predicted value of Y_i
N = Number of data points

The residuals are squared to ensure negative and positive values do not cancel when they are summed. Figure 1.17 illustrates the graphical relationship among the test data (Y_i), the model predictions ($Y_{i,\text{Pred}}$), and the residuals that appear in Equation (1.15). The best estimates of the model's dynamic parameters, e.g., K_P, τ, θ for a first-order model or K_P, τ_1, τ_2, τ_3 and θ for a second-order model, are calculated by a numerical optimization algorithm. Statistical calculations and "goodness of fit" criteria are used to estimate the uncertainties in the model. These first- and second-order dynamic parameter values are far better estimates than parameter values determined graphically, and they are used whenever possible.

If model predictions do not fit the data well, the model structure may be wrong (second-order model more suitable than first-order model) or the plant data collected is poor, e.g., step change magnitudes too small, data span too narrow, a new steady state was not reached, or the effects of disturbances are included in the data. But note that, in many cases, when the data fit is poor, a specific reason cannot be identified.

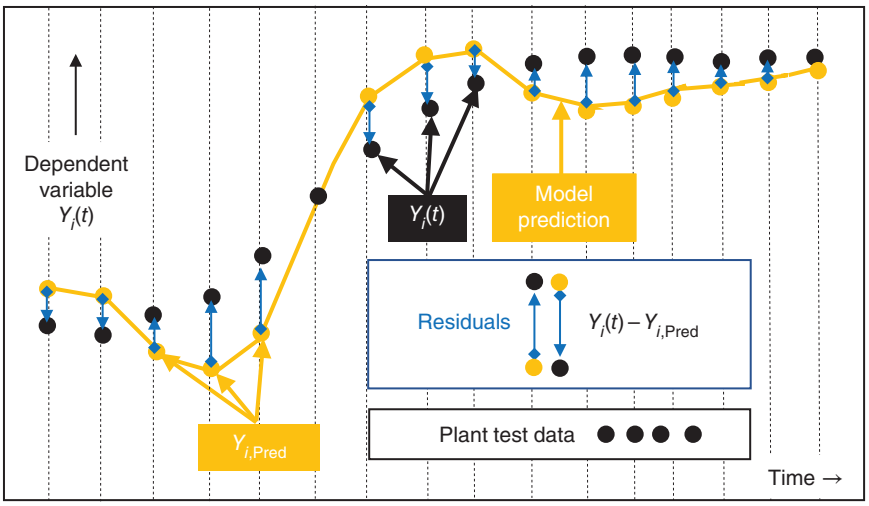


Figure 1.17 Plant Data, Model Prediction and Definition of Residual

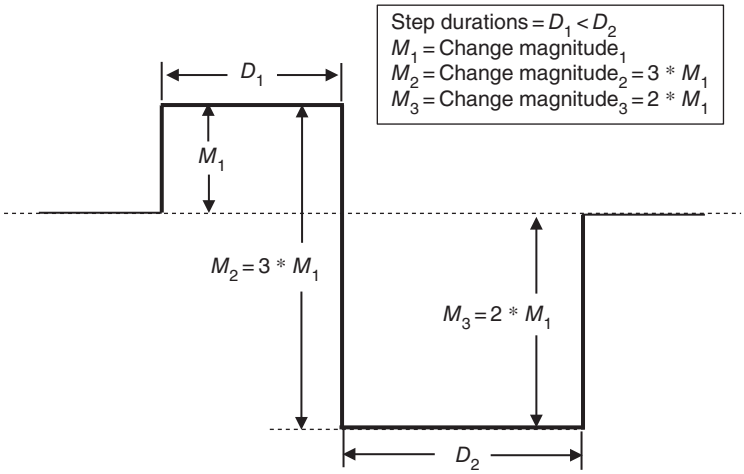


Figure 1.18 Independent Variable Rectangular Double Pulse Input Change

If a statistical curve fitting routine is used to determine the dynamic parameters, the following independent variable rectangular double pulse input change is desirable:

The independent variable's input change pattern in Figure 1.18 is advantageous for the following reasons:

- Both positive and negative step changes are executed; in part, this compensates for the effects of nonlinearities.
- Three different sized step changes are performed; this also compensates for the effects of nonlinearities.
- Step duration D_2 is chosen long enough for the dependent variable to approach and line out at a steady state value. This requires a reasonable initial estimate of the TTSS.
- The change magnitude M_1 is chosen large enough to overcome the negative consequences of noise.
- Since the independent variable returns to its initial steady state value, the dependent variable will also return close to or at its initial steady state value provided the system is close to linear.

1.7 Classification of Process Control Strategies

Process control strategies implemented in petrochemical plants can be ordered in a rational way (via *Levels*) based on control need (Ref. [9]). Control strategies at the lowest levels are the most critical, and control strategies at the higher levels less critical but more profitable. The most critical process control need is safety protection at Level 1, followed from lower to higher levels by basic regulation at Level 2, advanced regulation at Level 3, conventional control applications – supervisory controls at Level 4, and MVCC and online real-time optimization (RTO) at Level 5.

1.7.1 Control Strategy Hierarchy: Relative Costs, Resources, and Credits

The following table lists the five levels of process control strategies discussed in this section. The relative magnitudes of the hardware costs, manpower costs, potential credits, and reliability needs

are indicated for each level. Note that manpower costs are indicative of the control strategy's complexity and functionality requirements.

	Control Strategy	Hardware Costs	Manpower Costs	Potential Credits	Reliability Needs
5	Advanced Control Applications (MVCC and RTO)	Lowest	Highest	Highest	Lowest
4	Conventional Control Applications				
3	Advanced Regulatory Controls				
2	Basic Regulatory Controls				
1	Safety Protection	Highest	Lowest	Lowest	Highest

At Level 1, safety protection strategies are implemented in standalone devices independent of the DCS; they are critical and a prerequisite for operations to proceed. At Level 2, the basic regulatory controls provide the minimum control functionality that ensure operability, stability, and on-specification products. Safety protection and the basic regulatory controls have the highest reliability requirement and the highest hardware costs, but their maintenance manpower costs are relatively low.

At Level 3, the advanced regulatory controls have essentially the same control objectives as the basic regulatory controls, but their advanced features help improve control performance. Their hardware costs and credit potential are closer to the basic regulatory controls at Level 2, but their manpower costs and reliability needs are closer to conventional control applications at Level 4.

At Levels 4 and 5, conventional and advanced control application credit capture is the highest. These applications require complex and, in some cases, cutting-edge functionality; their manpower costs for implementation and maintenance are the highest. Their reliability is normally not a major issue because the advanced and basic regulatory controls (at Levels 2 and 3) run regardless of the advanced and conventional control application On/Off status. However, because their credit capture potential is usually quite high, periods of downtime will be reported, and issue resolution encouraged by plant management.

For the most part, process data, control statuses, and equipment statuses are pulled up from lower to higher levels, and calculated SPs and actuator positions (valve positions) are pushed down from higher to lower levels. SPs calculated by MVCC and RTO at Level 5, conventional control applications at Level 4, and advanced regulatory controls at Level 3 are stored to controllers at lower levels but *never* to a safety protection device at Level 1. The DCS never outputs to a safety protection device, and data from these devices (digital inputs and/or outputs) required for *control* in the DCS (Levels 2–5) are hardwired to DCS process connected devices. The information flow among advanced control applications (RTO and DMC), conventional advanced (supervisory) control applications, advanced regulatory controls, and basic regulatory controls is shown in Figure 1.19. Note that RTO applications almost always output to DMC *external targets*. However, RTO will on occasion (but infrequently) output a SP to a conventional advanced (supervisory) control application's CV.

Important: For control strategies at any level to meet their control objectives, control strategies at *lower* levels must be performing properly and achieving their control objectives.

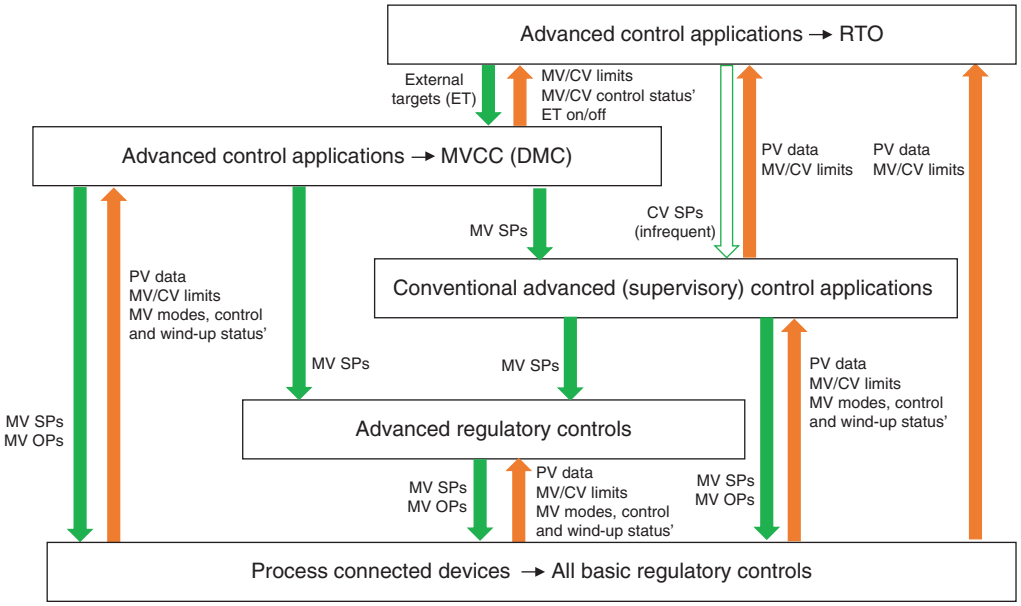


Figure 1.19 Information Flow in the Control Hierarchy

1.7.2 Safety Protection

At the lowest level is the hardware and software that provide safety protection, safeguarding equipment, the process units, and operating personnel. The safety protection system initiates automatic shutdown and start-up systems after warnings/alarms indicate potentially hazardous circumstances. They also prevent the start-up of equipment when applicable warnings/conditions are detected. These systems are implemented in standalone, multi-redundant process connected devices, Programmable Logic Controllers (PLCs) independent of the control system. The DCS never outputs to the PLC. PLC information (digital inputs and/or outputs) required for *control* are hard wired to DCS process connected devices; any direct connections to the DCS are for operator information only. Safety protection equipment is normally configured, programmed, and maintained by non-control engineer specialists.

Alarm systems are also safety protection devices. Alarms are prioritized according to their importance and potential impact on safety. Alarms that impact safety and require immediate operator action are assigned to the emergency category and are implemented in hardware, independent of the DCS. Emergency alarms are usually annunciated at the operator's console and backlit alarm windows. Other alarms are issued by process connected devices in the DCS or more complex control strategies in process connected devices. Although control engineers are involved in the planning and organization of the alarm management system, it is normally non-control engineers who are responsible for its configuration and maintenance.

1.7.3 Basic Regulation: Basic Regulatory Control

The sample control loops presented and discussed in Section 1.4.3 are all examples of *basic regulatory* controls, the next higher control level just above the safety protection level. They provide the minimum control functionality required to ensure operability, stability, and the production of on-specification products. All single control loops, e.g., flow, level, pressure, and temperature controllers and simple cascades such as Level → Flow, Temperature → Flow, Temperature → Pressure

are included in this classification. Basic regulatory control loops are *always* implemented in process connected devices that are redundant. They are typically configured and maintained by the instrument maintenance group, although control engineers support the tuning of troublesome loops – most often level and temperature controllers, and all control loops that have an impact on more complex, advanced control strategies.

1.7.4 Advanced Regulation: Advanced Regulatory Control

Basic regulatory control is enhanced by the addition of one or several control components such as ratio control, feedforward control, control of an important calculated process engineering PV (e.g., reboiler heat duty), etc. The enhanced control strategies that result are classified as *advanced regulatory* controls and reside in the next higher control level. They have the same control objectives as the basic regulatory controls, i.e., operability and stability, but the advanced features help improve control performance. An advanced regulatory control strategy example is presented next, but with much of the design details omitted because this control strategy is presented in Section 10.3.2, and all design issues and their resolutions are discussed there. In the discussion that follows, notice the prominent role that the process' configuration and operating objectives play in the development of the control strategy's design; this is an essential need for the design of most control strategies. This will be reinforced throughout the book when numerous industrially important, advanced regulatory control and conventional control application strategies are presented and discussed.

To illustrate the *advanced* regulatory features, first look at the *basic* regulatory controls for the bottom section of Deethanizer (DeC_2) tower T-209 in Figure 1.20. Tray temperature TC261, an inferential indicator of T-209 bottoms quality, is controlled by manipulating reboiler E-233's hot side bottom pumparound (BPA) bypass flow, via three-way bypass valve TV261, to either increase E-233 duty (reduce bypass flow, increase flow to E-233) or decrease E-233 duty (increase bypass flow, decrease flow to E-233). Temperature control via TC261 → TV261 is a basic regulatory control loop. *Note*: Maintaining TC261 constant in principle maintains bottoms C_2 content constant, but because TI261 is only an *inferential indicator* of bottoms quality, not necessarily at the *desired*

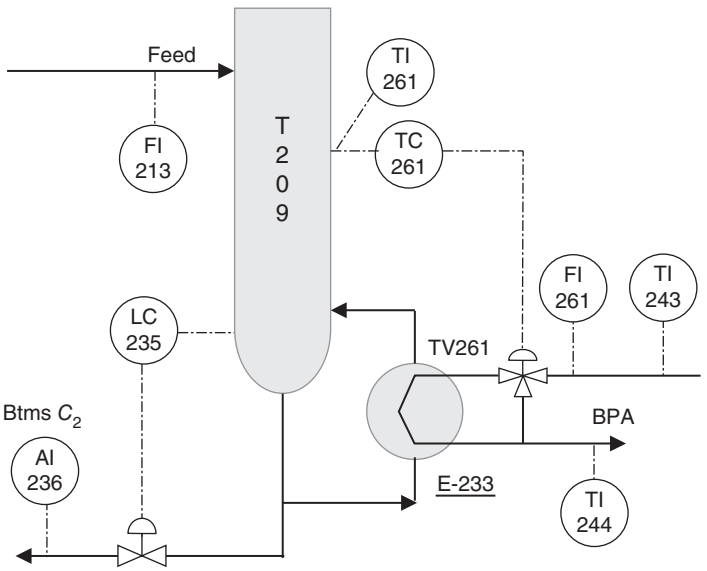


Figure 1.20 T-209 (DeC_2) Tray Temperature Control – Basic Regulatory Controls

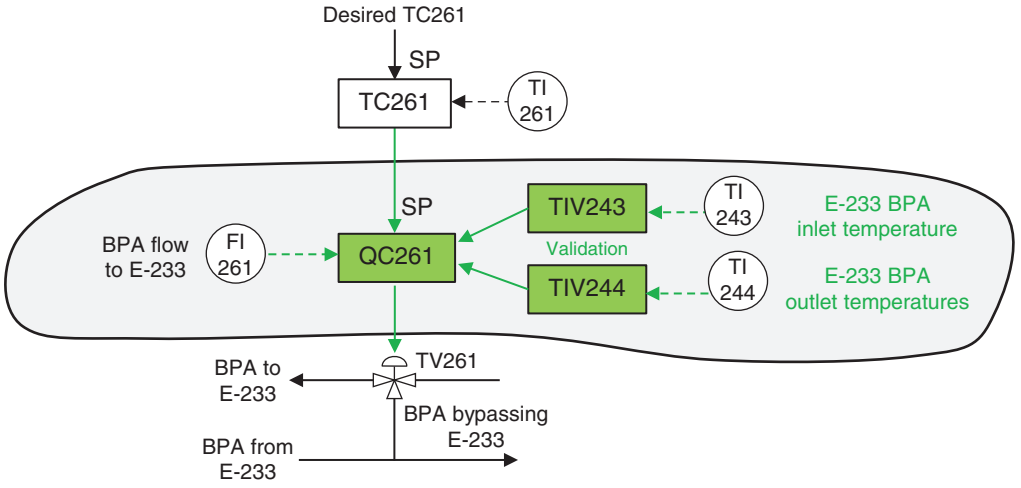


Figure 1.21 T-209 (DeC_2) Tray Temperature Control – With Reboiler Duty Control

value. Although not relevant to this discussion, with the bottoms C_2 content measurement available (AI236 PV), direct control of bottoms C_2 content is possible.

E-233 BPA flow (FI261) and inlet temperature (TI243), and T-209 feed rate (FI213) are all set by upstream or downstream unit operations. Changes in FI261 PV and TI243 PV directly affect E-233 duty and hence will upset tray temperature TC261 PV. Changes in FI213 PV will change the tower's L/V ratio, the tower's separation and thus tray compositions, and ultimately, they will also cause TC261 PV to deviate from TC261 SP. Therefore, upsets in any one of these three PVs will impact TC261 PV, and corrective control action starts *after* deviations in TC261 PV from TC261 SP are detected.

This is a classic example of where feedforward control for reboiler duty and feed rate upsets will improve control performance. By calculating and controlling E-233's duty and configuring TC261 to reset the duty controller's SP, corrective action for reboiler duty upsets will be initiated *before* TC261 PV deviates from TC261 SP. Figure 1.21 illustrates how the basic regulatory control strategy in Figure 1.20 can be upgraded by adding reboiler duty control (QC261).

The advanced regulatory control components added to the basic TC261 regulatory control strategy are enclosed in the shaded area and shown with green background. Because QC261 outputs to a valve, and the relationship between heat duty and valve position is *not* linear, QC261 must be configured as a PID controller; its PV is the calculated duty:

$$\text{QC261 PV} = \text{Scale} * \text{FI261 PV} * C_P(\text{BPA}) * (\text{TIV243 PV} - \text{TIV244 PV}) \tag{1.16}$$

E-233 heat duty QC261 PV is calculated using hot side inlet (TI243) and outlet (TI244) temperatures, and flow rate (FI261). TI243 and TI244 are both DTI inputs (externally multiplexed) and not analog inputs (4–20 mA); they must be validated[13] (in TIV243 and TIV244) before they are used in the duty calculation. C_P is the BPA's heat capacity, assumed constant at a process technical services (PTS) specified value. *Scale* is a unit conversion factor that ensures QC261 PV has the desired EUs.

Figure 1.22 illustrates how the advanced regulatory control strategy in Figure 1.21 is further enhanced by adding feed rate feedforward control (FCG213) and heat duty to feed ratio control (RC261):

13 See Chapter 8, Section 8.8.

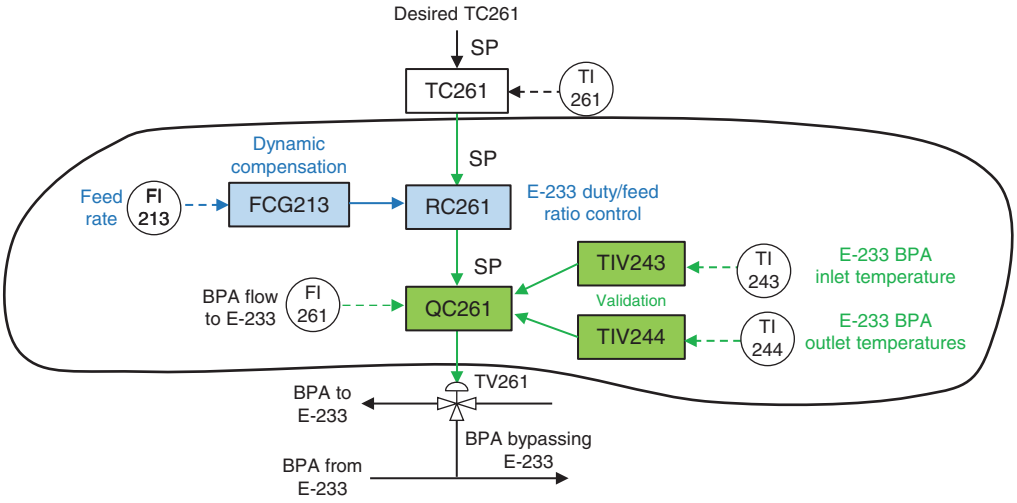


Figure 1.22 T-209 (DeC_2) Tray Temperature Control – Reboiler Duty and Feed Rate Feedforward Control

These additional control components, e.g., feed rate feedforward and duty/feed ratio control, are now also enclosed in the shaded area, this time shown with blue background. RC261 is the heat duty/feed ratio controller whose PV is the calculated ratio:

$$\text{RC261 PV} = \text{Scale} * \frac{\text{QC261 PV}}{\text{FI213 PV}} \tag{1.17}$$

FI213 PV is the measured feed rate and QC261 PV the calculated reboiler heat duty. *Scale* is another unit conversion factor that ensures RC261 PV has the desired EUs.

RC261 can be configured as a standard PID controller, but normally it is configured with a ratio control algorithm[14] that performs an arithmetic control calculation:

$$\text{RC261 OP} \equiv \text{QC261 SP} = \text{RC261 SP} * \text{FCG213 PV} \tag{1.18}$$

Note that RC261 PV has no impact on RC261's control calculation in Equation (1.18) – it is RC261's "wild flow" input, FCG213 PV, the *dynamically compensated* feed rate, that RC261 SP multiplies in the control calculation. Dynamic compensation of the feed rate is required when the tray temperature's response dynamics (K_P, θ, and τ) to a heat duty change are visibly different from the tray temperature's response to a feed rate change.[15] The dynamic parameters of the two responses, e.g., their K_P, θ, and τ values, are estimated using plant test data and are the feedforward controller's tuning constants.

Ratioing heat duty to feed is intuitively correct if, in the current operating region, the process is close to linear. If so, the reboiler's heat input requirement should be proportional to the feed rate – e.g., if the feed rate goes up by 10%, the heat duty required to maintain the tray temperature constant is expected to increase by roughly 10% as well.

Like almost all ratio feed rate feedforward and reboiler duty controls, there are several especially important "what if" (e.g., measurement signal failures), initialization and wind-up issues that must be managed by the configurations of FCG213, RC261, and QC261. For this specific control strategy, these details are presented in Section 10.3.2. There are numerous other applications presented and discussed in this book that include ratio feed rate feedforward and duty controls; all require these same safeguards for proper control performance.

14 See Chapter 2, Section 2.6.
15 See Chapter 3, Section 3.3.

1.7.5 Supervisory Control: Conventional Advanced Control Applications

Conventional advanced control applications, historically called supervisory controls, calculate and output SPs to both advanced regulatory and basic regulatory controllers. They provide both PID and model-based control of product quality, improve energy utilization, move operations closer to and maintain them at constraints, and meet reactor conversion targets. A conventional control application example is presented next, but again with much of the design details omitted because this control strategy is presented in Section 10.2.4, and all design issues and their resolutions are discussed there. Once again notice the prominent role that the process' configuration and operating objectives play in the development of the control strategy's design. And note again that many other industrially important conventional control applications are presented and discussed in this book.

To illustrate how a conventional control application's design structure evolves, consider the basic regulatory controls for the hydrodesulfurization (HDS) Product Stripper T-635 in Figure 1.23. The HDS process is a fixed bed catalytic reaction system that removes feed sulfur, nitrogen, metals and Conradson carbon and produces fluid catalytic cracking unit (FCCU) feed stocks and desulfurized heavy fuel oils. The reactor effluent is separated into liquid and gas products, and the liquid products are heated and fed to a product stripper such as T-635 in Figure 1.23. Tail gas (uncondensed overhead vapor), wild naphtha (distillate), wild gas oil sidestream (FCCU feed), and heavy fuel oil (bottoms) products are separated.

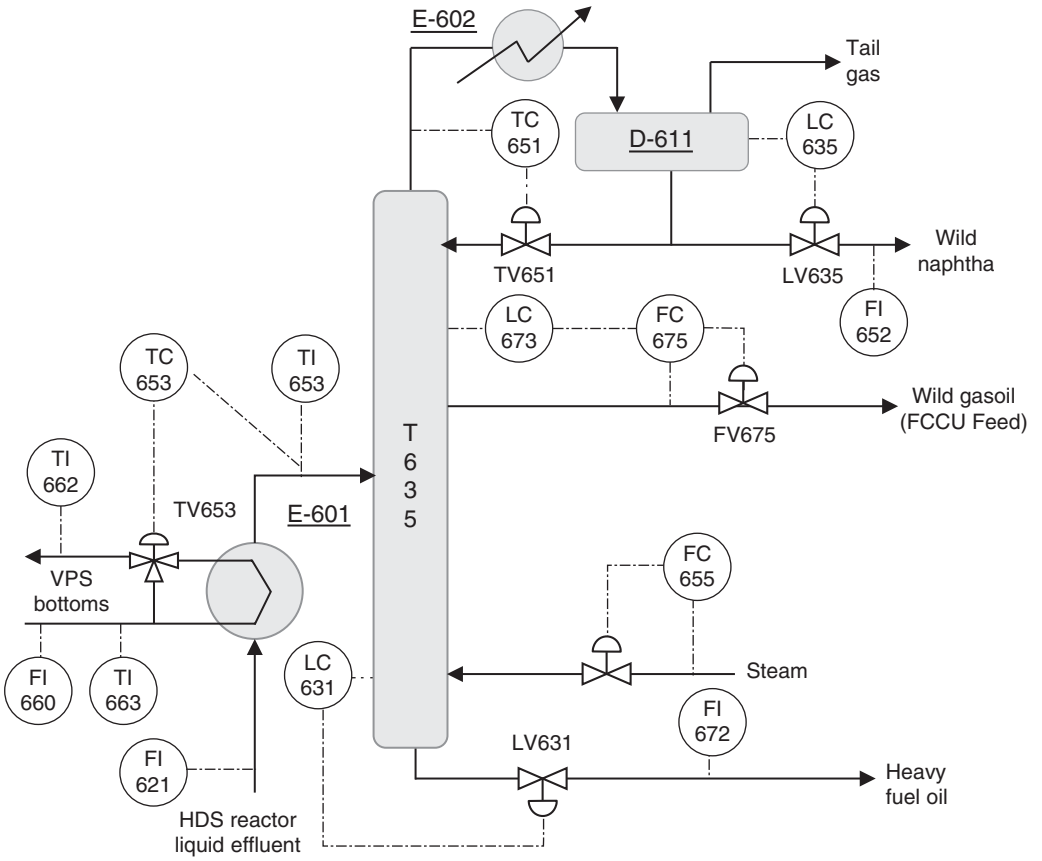


Figure 1.23 Product Stripper Tower T-635 Configuration and Feed Preheat Control

After H_2S removal, the tail gas is sent to the treat gas and/or fuel gas systems. The wild naphtha becomes either catalytic reformer feed or MOGAS blend stock. The wild gas oil product is a FCCU feedstock. Preheat exchanger E-601 supplies the heat required to vaporize the components removed as liquid products *above* the feed tray, and the liquid returned as reflux to insure good fractionation above the sidestream draw-off. The higher the preheat duty, e.g., the higher the feed temperature, the more vapor is sent overhead and thus the more reflux, wild naphtha and wild gas oil produced. The overhead condenser E-602 capacity is sized to condense almost all the vapor produced in E-601 and not condensed by the tower's internal reflux. Steam stripping below the feed tray removes lower boiling (lighter component) hydrocarbons from the unvaporized feed, upgrading small amounts of heavy fuel oil to wild gas oil.

There are five key T-635 basic regulatory control loops: (i) overhead temperature control (TC651) manipulating reflux flow (via TV651) to set wild naphtha product (back-end) quality, (ii) reflux drum D-611 level control (LC635) manipulating wild naphtha product flow (FI652) via LV635 which sets wild naphtha yield, (iii) complete draw-off tray level control (LC673) manipulating sidestream draw off flow (FC675 SP) to set wild gas oil product yield, (iv) level control (LC631) manipulating bottoms product flow (FI672) via LV631 which sets heavy fuel oil yield, and (v) temperature control (TC653), manipulating feed preheat exchanger E-601 hot side (VPS bottoms) bypass flow (via three-way bypass valve TV653), to set T-635 feed temperature.

The liquid hold-up on the draw-off tray, and thus the wild gas oil product yield, varies with the amount of vapor produced in preheat exchanger E-601 (E-601 duty). TC653 SP determines the feed preheat duty and thus the tower's vapor loading. Because overhead temperature control (TC651) manipulates reflux flow, E-601 duty also affects the tower's liquid loading. Since the wild gas oil is normally FCCU feed, the operating objective is to maximize gas oil yield *by maximizing E-601 duty*, i.e., maximizing TC653 SP, as limited by tower operating constraints. Operating experience indicates the most likely constraints to limit feed preheat are a full open reflux valve (TV651) and/or a full open distillate valve (TV635); neither wild gas oil draw-off valve (FV675) or heavy fuel oil flow valve (LV631) becomes an active constraint *before* either the reflux or distillate flow valve becomes limited. Clearly, to meet the wild gas oil yield maximization operating objective, TC653 SP must be increased. However, TC653 SP cannot exceed its guideline maximum value specified by the PTS group, and the reflux and distillate flow valve positions must stay in controlling range.

The control of T-635 feed temperature and E-601 duty can be enhanced by adding the same control components, e.g., reboiler duty control, feed rate feedforward control, and duty to feed ratio control, added to Figures 1.21 and 1.22 in Section 1.7.4. These additions, and configuring TC653 to reset the duty/feed ratio controller's SP will initiate corrective actions for feed rate and E-601 hot side inlet temperature and flow upsets *before* TC653 PV deviates from TC653 SP. Figure 1.24 illustrates how the basic regulatory control strategies in Figure 1.23 can be upgraded by adding reboiler duty control (QC653), feed rate feedforward control (FCG621), and duty/feed ratio control (RC653) – all additions are enclosed in the shaded area, and shown in green. For this control strategy, TC653 SP is an operator entered desired T-635 feed temperature.

E-601 heat duty controller QC653 outputs to a valve so it must be configured as a PID controller. Its PV is calculated using Equation (1.16) written in terms of QC653 PV, E-601 hot side inlet (TIV663 PV) and outlet (TIV662 PV) temperatures, and E-601 hot side flow FI660 PV. The heat duty/feed ratio PV (RC653 PV) is calculated using Equation (1.17) written in terms of duty QC653 PV and feed rate FI621 PV. RC653 is configured with a ratio control algorithm; its control output is calculated using Equation (1.18) written in terms of RC653 SP, the dynamically compensated feed rate FCG621 PV, and E-601 heat duty QC653 SP.

To achieve the tower's primary operating objective – to maximize wild gas oil yield – the control strategy in Figure 1.24 requires operator action and attention, e.g., to increase TC653 SP and

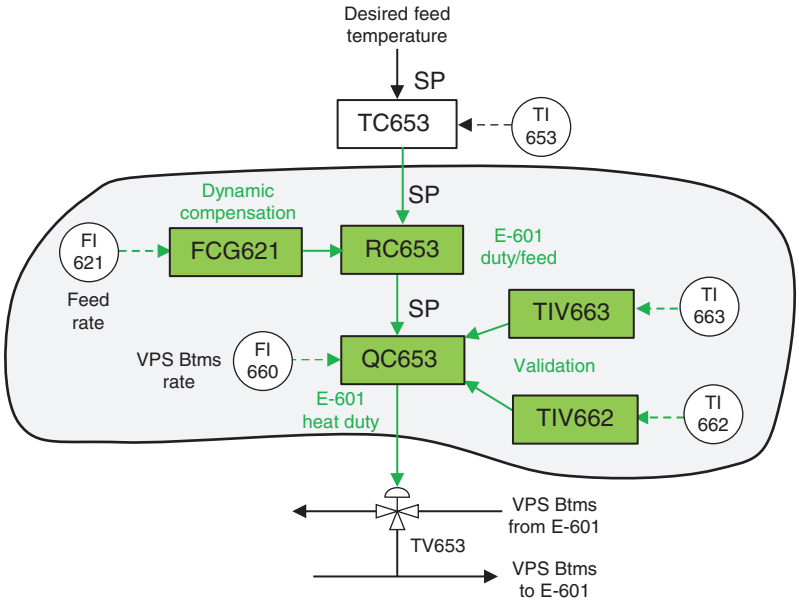


Figure 1.24 T-635 Feed Preheat Control – Reboiler Duty, Duty/Feed, Feed Rate Feedforward Control

monitor reflux and distillate flow valve positions. Even if the incentives are high, operators will normally give a "constraint pushing" strategy a lower priority than those control strategies that stabilize operations, confine operations to safe operating regions, or meet product quality targets.

The control strategy in Figure 1.24 can be improved to automate the "constraint pushing" part of the strategy. The enhanced design, shown in Figure 1.25, converts the advanced regulatory control strategy in Figure 1.24 to a typical supervisory control application, i.e., conventional constraint control[16] with three constraint variables (T-635 feed temperature, reflux, and distillate valve positions) and a single MV (E-601 three-way bypass valve position, TV653). The supervisory control additions – reflux valve position controller VTV651, distillate valve position controller VLV635, and the low signal selector (LSS) – are enclosed in the shaded area and shown with green background, connectors and text. Each constraint variable has its own dedicated PID controller (QC653, VTV651, and VLV635) trying to adjust the one MV (TV653) to drive its PV value to its constraint limit, i.e., its PID's SP (QC653 SP – as calculated by RC653, to drive TC653 PV to TC653 SP, VTV651 SP, and VLV635 SP). Note that for this control strategy, TC653 SP is the PTS group's desired maximum feed temperature guideline value. Both valve position constraints are also maximum constraints.

With no constraint variable violating, an increase in preheat duty (MV = TV653) will drive all three constraint variables toward their constraint limit. The *lowest* amount of preheat (after the increase) will in principle drive the constraint variable closest to its limit (the *active constraint*) to its limiting value without driving the other two constraint variables into violation. With all constraint variables violated, the lowest amount of preheat (after the decrease) will drive the active constraint (constraint variable farthest from its limit) to its constraint limit while moving the other two constraint variables out of, and away from, violation. Likewise, the same logic can be applied to cases in which either one or two constraint variables are violated – the lowest amount of preheat (after the decrease) will drive the active constraint variable to its limiting value without forcing any additional violations. Thus, the three constraint variable PIDs in Figure 1.25 (QC653, VTV651, and VLV635)

16 Constraint control is discussed in detail in Chapter 7.

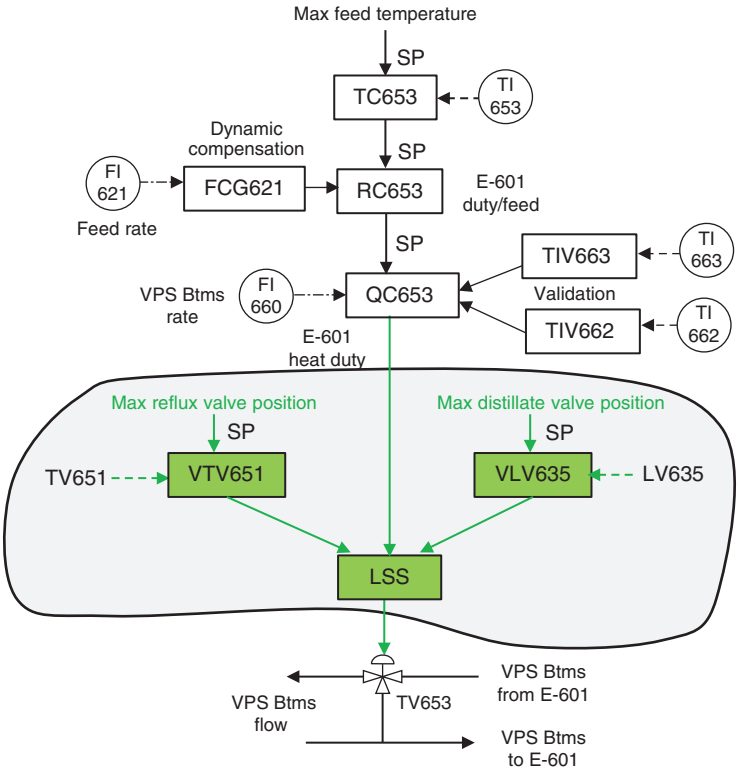


Figure 1.25 T-635 Feed Preheat Maximization

must output to a LSS, which chooses the smallest of the three PID outputs (the lowest preheat duty) to send to TV653.[17] This either satisfies the PTS group's *maximum* feed temperature guideline value (the max TC653 SP entered) via the TC653 → RC653 → QC563 cascade's control action, or limits E-601 duty and thus TC653 PV to ensure both valve positions remain in controlling range.

Like the control strategy in Figure 1.22, there are several especially important "what if" (e.g., measurement signal failures), initialization and wind-up issues that must be managed by proper configuration of FCG621, RC653, QC653, VTV651, VLV635 and the LSS. Remember the control strategy in Figure 1.25 is presented and discussed in detail in Section 10.2.5, where these configuration features, as applied to this application, are explained. Also, a detailed discussion of conventional constraint control design issues can be found in Section 7.8. Finally, there are numerous other applications presented and discussed in the book that include conventional constraint control, and all require these same important safeguards to achieve satisfactory control performance.

1.7.6 MVCC and RTO: Advanced Control Applications

Advanced control applications, e.g., MVCC and RTO applications, are far more complex than the advanced regulatory and conventional advanced control applications discussed to this point.

17 As will be explained in Sections 7.3.4 and 7.4.4, criteria for choosing the selector (LSS or HSS) depend upon the effect an increase in the MV has on all constraint variables – specifically, if an MV increase drives all constraint variables towards violation, an LSS is the correct choice. Thus, for the strategy in Figure 1.25, this general criterion provides the same answer as the analysis presented here.

However, they are also *far more profitable* and capture a good share of the credits associated with control applications.

MVCC applications such as DMC calculate and output multiple SPs to underlying controllers resident at lower hierarchical levels – as indicated in Figure 1.19, to conventional advanced control applications, advanced regulatory and basic regulatory control strategies. In many cases, DMC controllers replace some conventional control and advanced regulatory control applications by incorporating their CVs and MVs into the DMC's controller's scope. DMC controllers hold the process within or at constraint limits, as specified by the site's operating plan and entered by operators, although that operation *may or may not be optimal. Remember*: For a DMC controller to achieve its control objectives, its underlying conventional control applications and advanced and basic regulatory control strategies must be on control and performing effectively.

RTO applications calculate the process conditions that either maximize or minimize a carefully developed *economic* objective function, e.g., RTO either maximizes profit or minimizes total costs. RTO applications provide optimal SPs for DMC CVs in one or more underlying DMC controllers; these RTO calculated SPs are called *external targets* in DMC. For an RTO application to achieve its optimization objectives, its underlying DMC controllers must be on control and must be able to move plant operations to the optimal operating points calculated by RTO.

DMC controllers incorporate linear, dynamic models that quantify the dynamic relationships among CVs and the MVs, and DVs that impact the CVs. DMC drives multiple constraint variables (the CVs) to maximum or minimum constraint limits by simultaneously adjusting multiple MVs. DMC is thus a truly multivariable controller that accounts for the real-world certainty that many important CVs are affected by more than one MV and that some MVs impact multiple CVs. A critical step in the DMC control calculation, as it is for all model-based controllers, is the update of the DMC model. Prior to each control execution, DMC models are updated online in real time by calculating CV *prediction errors*, which are then used to shift the model's CV predictions to match plant data at the current operating point.

RTO applications incorporate complex, *nonlinear*, first-principle, *steady state* models that are solved as part of the optimization calculation. The model equations are updated online in real time prior to each execution pass – some of the steady state model's physical parameters (e.g., a mass transfer coefficient) are configured to be back calculated to match the model predictions to plant data at the current operating point. Because RTO models are valid *only* at steady state, the RTO application's calculation sequence includes a *steady-state detection* algorithm. An RTO application's optimization calculation is initiated *only* if the steady state detection algorithm declares the process is at steady state.

Thus, the RTO application executes *asynchronously* – it is not scheduled to execute at a *specific* frequency, but executes only when the process is declared at steady state. RTO execution frequencies range from every hour to every four or more hours. Remember, an RTO application's underlying DMC controllers must be on control and performing effectively. This is important for many reasons, but it is especially important for an RTO application that can execute *only* if the process is declared to be at steady state. Therefore, DMC must not only move plant operations to the optimal operating point calculated by RTO on its *last* execution pass, it must hold the plant steady there. If not, steady state detection may fail, and if so, a new RTO optimization calculation cannot start.

Conventional control applications and advanced regulatory controls are the focus of the many industrially important control strategies presented and discussed in detail in this book. The complexity of the advanced control application technology underlying both DMC and RTO is well beyond this focus, and neither DMC nor RTO applications are included. However, for many of the conventional control applications discussed, process scenarios that suggest a potential need for either DMC or RTO are identified and discussed. One example of such a conventional control

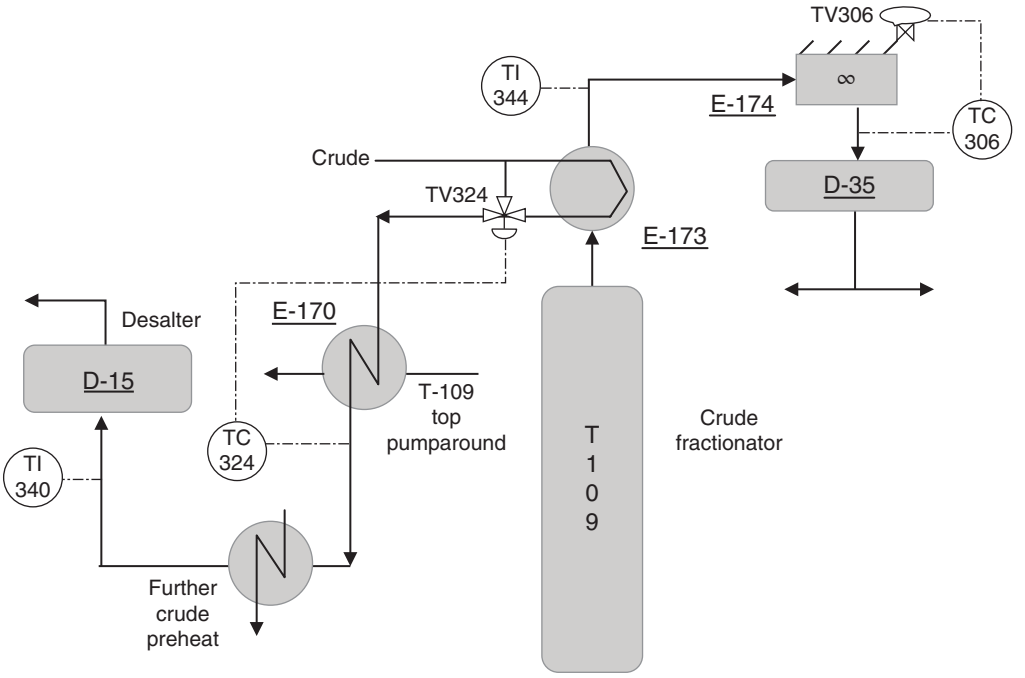


Figure 1.26 Fractionator T-109 Feed Preheat Configuration Upstream of Crude Desalter

application is presented next. Because this conventional control strategy is presented and discussed in Chapter 11 (Energy Conservation Controls) – Section 11.1.1, where design features and issues are discussed, most of the design details are omitted here and the focus is on the potential need for DMC.

Figure 1.26 shows the portion of a crude fractionator's (T-109) preheat train upstream of the crude desalter (D-15). Crude flows first to overhead condenser E-173, then to T-109 TPA exchanger E-170, and to other exchangers before flowing to Desalter D-15. An increase in E-170 cold side (crude) outlet temperature controller's SP (TC324 SP) will reduce crude bypass in E-173 (via three-way bypass valve TV324), and thus increase heat recovery to crude in E-173 and reduce heat lost to air in E-174. Thus, an operating objective is to maximize heat recovery to crude in E-173 and to minimize heat lost to air in E-174 by maximizing E-170 crude outlet temperature (TC324 SP).

Crude exiting E-170 is heated further in other downstream exchangers before entering D-15. Crude desalting is required to minimize downstream fouling, corrosion, and acid formation. Desalter efficiency can be limited by high temperatures; for D-15, there is an operating temperature above which desalter efficiency is too low to ensure sufficient crude desalting. Thus, a second operating objective is to maintain D-15's inlet temperature (TI340 PV) below its maximum to ensure acceptable desalting.

As E-173 duty increases, T-109 overhead vapor temperature (TI344 PV) decreases. However, TI344 PV must remain above its dew point to minimize corrosion in the overhead line. Also, as TI344 PV decreases, E-174 outlet temperature tends to decrease, and temperature controller TC306 reduces E-174 duty to maintain TC306 PV at its SP. Therefore, two other operating objectives are to maintain TI344 PV above its minimum value to avoid corrosion in the overhead line, and to maintain E-174 outlet temperature controller TC306 in controlling range (TC306 OP above its minimum value).

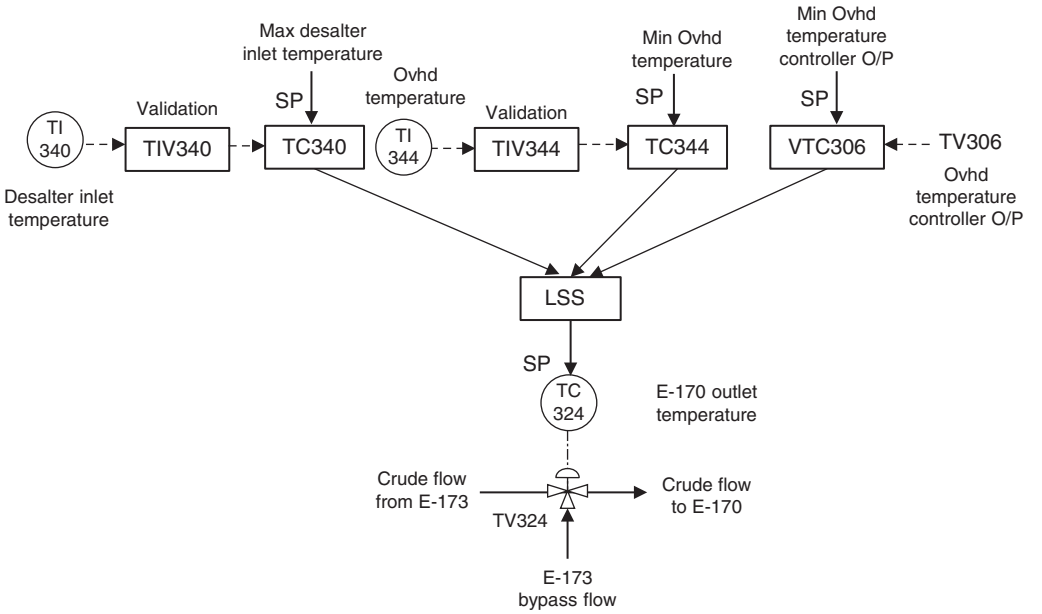


Figure 1.27 Maximization of Heat Recovery to Crude in T-109 Overhead Condenser E-173

To summarize, the control strategy must maximize E-170 cold side outlet temperature (TC324 PV) while satisfying maximum desalter inlet temperature (TI340 PV), minimum overhead temperature (TI344 PV), and minimum TC306 control output (to maintain TC306 OP in controlling range) constraints. The conventional constraint control strategy in Figure 1.27 achieves these control objectives. The strategy's design intent can be summarized as follows: By increasing TC324 SP until one of the three constraint variables reaches its constraint limit, heat recovery to crude in E-173 can be maximized.[18] Remember, the discussion of design details for the strategy in Figure 1.27 is deferred to Section 11.1.1 so that the focus here can be on process conditions that could enlarge the scope of this control strategy and suggest the need for DMC or even RTO.

There are usually multiple heat exchangers in a fractionator's pumparound circuit. However, in Figure 1.26, other than E-170, no other information is provided about T-109's TPA configuration. The constraint control strategy in Figure 1.27 assumes that there are no adverse downstream process or economic consequences for TPA operations as E-170 hot side (TPA) outlet temperature is reduced to maximize heat recovery to crude in E-173. However, assume a TPA exchanger downstream of E-170 becomes constrained and is no longer able to satisfy its heat duty requirement, for example, providing reboiler heat to meet a light-ends tower's tray temperature control objective. If so, heat recovery to crude in E-173 may need to be constrained further to maintain the light-ends tower's tray temperature controller's output in controlling range. And the range chosen, e.g., output less than 75% versus less than 85%, may impact other tower variables and thus further enlarge the scope of the control strategy.

Although the control adjustments made to maximize heat recovery to crude in E-173 do not affect TPA draw-off flow or temperature, the control strategy also assumes that TPA return temperature is controlled and always remains at SP. A pumparound return temperature is almost always

18 In Figure 1.27, an LSS is the selector choice because an increase in TC324 SP (the MV) drives all three constraint variables toward their constraint limits. See discussions in Sections 7.3.4 and 7.4.4.

controlled by manipulating a pumparound heat exchanger's duty. Note that if TPA return temperature is not held at SP, even with constant TPA draw off flow and temperature, TPA duty is not constant and T-109 fractionation may be affected. Even though these fractionation upsets may not have a significant impact on product yields, constant pumparound return temperature is almost always the preferred operating strategy. For TPA return temperature to remain at SP, the return temperature controller's output must remain in controlling range. If this is not always true, heat recovery to crude in E-173 may also need to be constrained further to maintain the return temperature controller's output in controlling range.

There is also no information provided about the hot side process streams that exchanges heat with crude in heat exchangers downstream of E-170. Here, again, it is necessary to ensure that any potential constraints that can impact heat recovery in E-173 are identified and considered in the control strategy's design. Note that process upsets in these exchangers may affect desalter inlet temperature TI340 PV directly, and the E-173 heat recovery to crude control strategy must respond effectively to these disturbances. These interactions may also increase the control strategy's scope and impact the heat recovery to crude possible in E-173.

This discussion illustrates how important it is to ensure all potential constraints are identified and included in a control strategy's scope – in fact, one of the first tasks in control strategy design is to correctly define the strategy's scope. This is important because as the number of control and/or constraint variables increase, and even more importantly, if multiple MVs have significant control and/or constraint variable sensitivities, a multivariable constraint controller such as DMC is by far the best choice.

To look at an example where RTO may be needed, refer to Figure 1.23 in Section 1.7.5, which presents the HDS product stripper's (T-635) process configuration. As discussed there, the wild gas oil, an FCCU feedstock, is the most important product. The operating objective is to maximize wild gas oil yield by maximizing feed preheat. The maximization is subject to a maximum T-635 feed temperature (TC653) and maintaining the reflux (TV651) and distillate (LV635) flow valve positions in controlling range.

To maximize the wild gas oil product yield, reboiler E-601 duty and the heat lost to cooling water in overhead condenser E-602 are both maximized. Thus, for current operations, the energy-yield trade-off must favor yield improvement, i.e., the incremental value of a wild gas oil yield increase is greater than the incremental cost of the heat utilized and/or lost. However, there are many process unit and plant operations in which the trade-off does not have a consistent winner, and numerous competing energy-yield trade-offs must be resolved. For example, VPS bottoms, E-601's hot side fluid, no doubt exchanges heat with other upstream or downstream units, and its ability to supply the heat needed by its other energy consumers is impacted by E-601 heat duty. So not only is E-601 heat duty traded off versus T-635 wild gas oil yield, it may also need to be traded off against other VPS bottoms heat utilization needs. Hence the scope of the energy yield trade-off may be wider than T-635 and include upstream and downstream units. If real-time assessments of energy-yield tradeoffs are needed, RTO is the best choice.

1.8 Benefits of Control Applications

Control applications improve process control performance by reducing CV variability. Control application benefits are captured when operations are moved closer to a product specification or constraint limit to exploit the reduced variability. This requires changing an operating target, e.g., the CV's SP; unless the SP is changed and the operation is moved closer to the specification or constraint limit, the bulk of the application's potential benefits are lost.

To illustrate the basis for estimating the benefits of control applications, refer to the discussion of product stripper T-635 in Section 1.7.5, e.g., its basic regulatory control (see Figure 1.23), advanced regulatory control (see Figure 1.24), and conventional constraint control application (see Figure 1.25). T-635's operating objective is to maximize its sidestream's yield (wild gas oil) by increasing T-635 feed inlet temperature (TC653 SP). Feed inlet temperature cannot exceed the maximum allowed value specified by the PTS group. TC653 manipulates preheat exchanger E-601 duty by adjusting E-601 feed bypass valve TV653. As TC653 SP, and hence E-601's duty increase, reflux and distillate valves are driven toward their maximum allowed openings. Thus, increases in TC653 may be limited by the need to maintain both valves in controlling range.

Figure 1.24 illustrates how the basic regulatory control strategy (TC653 → TV653) in Figure 1.23 is upgraded by adding three advanced regulatory control components, e.g., E-601 duty control (QC653), feed rate feedforward control (FCG621), and duty/feed ratio control (RC653). These additions enable corrective actions for feed rate and E-601 hot side inlet temperature and flow upsets to begin *before* TC653 PV deviates from TC653 SP. These feedforward enhancements improve control performance, and feed temperature (TC653 PV) variability about its SP (TC653 SP) is reduced. The reduced variability is an enabler for operators to increase TC653 SP and to move TC653 closer to its PTS specified maximum target.

The control strategy in Figure 1.24 requires operator actions and attention, e.g., increasing (or decreasing) TC653 SP, and reflux and distillate flow valve position constraint monitoring. The control strategy in Figure 1.24 can be improved to automate the "constraint pushing" part of the strategy; the enhanced design in Figure 1.25 converts the advanced regulatory control strategy in Figure 1.24 to a typical conventional advanced control application, i.e., constraint control. Each constraint variable has its own dedicated PID controller (QC653, VTV651, and VLV635) whose outputs compete to adjust the one MV (E-601 three-way bypass valve TV653) to drive its PV to its constraint limit, i.e., TC653 SP, the PTS group's desired *maximum* allowable feed temperature guideline value, and VTV651 SP and VLV635 SP, reflux and distillate valve position maximum opening constraints.

Figure 1.28 is an illustration of how implementing the T-635 advanced regulatory control and conventional constraint control applications discussed above will capture control benefits. With only basic regulatory control (single loop control, TC653 → TV653), TC653 PV shows considerable variation about its SP. As a result, to ensure no violations, operators enter TC653 SPs far away from the PTS guideline max value. The difference between the actual SP and the PTS guideline value represents a lost opportunity, i.e., the yield improvement credits if the feed temperature is raised to the guideline maximum value, less the incremental cost of energy (if any); this is indicated as *Lost Opportunity #1* on Figure 1.28.

With the implementation of the advanced regulatory control application (Figure 1.24), TC653 PV variations about its SP are markedly reduced. With operators increasing TC653 SP over the time period shown in Figure 1.28, a portion of *Lost Opportunity #1*, identified during basic regulatory control operations, is recovered; the credit basis for the portion recovered is indicated as *Credit Capture Basis #1* in Figure 1.28.

The conventional constraint control application automates the "constraint pushing" part of the strategy. Instead of depending upon operators to increase TC653 SP, it drives TC653 up to the max TC653 SP PTS guideline value unless or until limited by either the reflux or distillate flow valve max opening constraint. Because some residual variations in TC653 PV remain, the temperature SP entered in the constraint controller is typically somewhat below the max PTS guideline value to ensure the guideline is not violated. As illustrated in Figure 1.28, a large portion of *Lost Opportunity #2*, identified during advanced regulatory control operations, is recovered by implementing constraint control. The basis for the portion recovered by constraint control is indicated as *Credit Capture Basis #2*, and *Lost Opportunity #3* remains as the incentive to further increase TC653 SP.

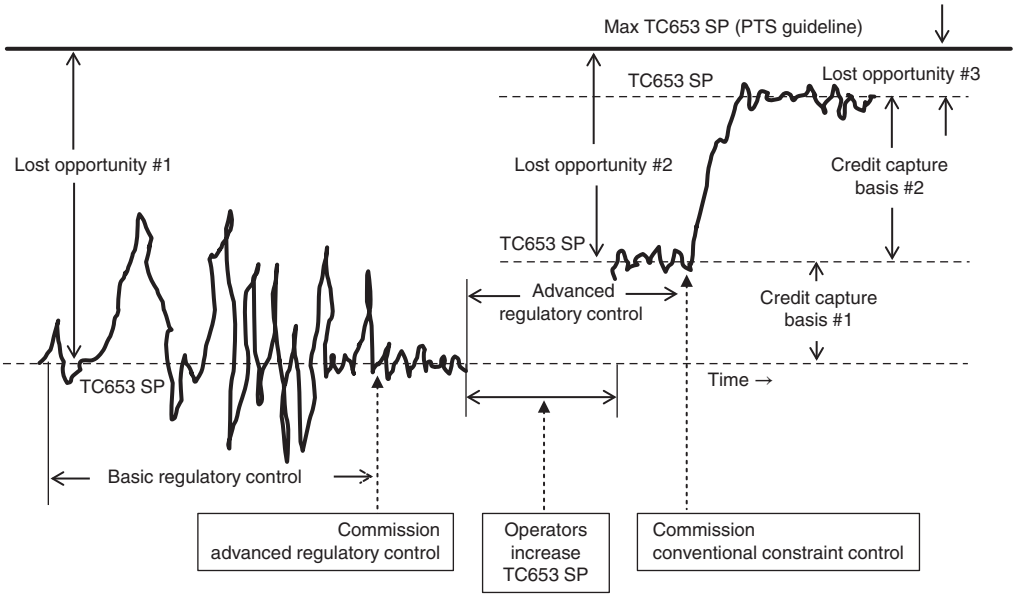


Figure 1.28 Basis for Control Application Benefits

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