The Impact of Control Loop Performance on Process Profitability

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Abstract

Strong emphasis is being placed on process optimization in the process industries. In many instances, this has led to large investments in control room equipment, supervisory control systems, modeling of plant processes, and advanced control algorithms on profit critical units of a production facility. While these investments have resulted in some overall improvements, the maximum return on investment is severely limited if the field control loop components are not properly matched to the control room system. It has been determined that small improvements in the performance of the automatic control system (the final control element, the transmitter, and the controller) will significantly impact the profitability realized in process optimization.

This paper will examine the impact that the control loop has on process optimization. Specific examples of process improvement will be presented. The use of proper control loop equipment analysis, as part of the optimization process, will be examined. In addition, the importance of various performance and design parameters of control valve assemblies will be explored.

Introduction

Inherent in the concept of process optimization is the reduction of process variability. Maintaining the process variable as close as possible to the set point will allow the process to operate closer to the plant's constraints and can have a tremendous impact on the bottom-line results due to increased throughput, yield, and quality, as well as a reduction in waste. Strong emphasis is being placed on process optimization in most of the process industries. Large investments in distributed control systems (DCS), supervisory control, and advanced control have been made in pursuit of process optimization and better profitability. The promise of improved economic return is a key driver for these investments.

Supervisory control and advanced control strategies have been considered important steps in achieving process optimization. However, if process optimization looks only at the control room equipment and software and does not consider loop field hardware performance, the true benefits of the investment will not be realized. Little can be gained by developing a sophisticated control room architecture that is capable of performing to half percent accuracy and then implementing that control strategy with a final control element that may only be capable of five percent accuracy at best!

Not only will poor control loop instrumentation prevent the process from achieving its full potential, but it can even aggravate the process control performance. M.J. Oglesby[1] states, "No amount of advanced control which relies on the use of poor field instrumentation can be expected to yield worthwhile benefits. Thinking of control as a hierarchy, everything must work well at the lower levels for the higher levels above to work." Bill Bialkowski, President of EnTech Control Engineering (Toronto), echoed this position when he stated that 80 percent of the control loops his firm has audited over the years actually fail to reduce process variability to an acceptable degree, and that poor dynamic performance of control valves was the major contributor to the problem[2].

In a comprehensive benchmarking study performed by Monsanto and eleven other chemical manufacturers it was determined that final control element performance had the largest impact on the cost of goods sold (COGS). As much as 1.5% of the COGS could be saved by employing best practices on the final control loop. Indeed, even the advantages promised from advanced regulatory control were limited if final control loop performance was not addressed.

In over 4,000 process loops audited by Fisher Controls, the performance of over half of the loops could be significantly improved if work was done on the final control element (valve, actuator, and/or positioner-I/P). In many of these audits, advanced control strategies had been used as the first step in process optimization. Without exception, we have found that significant improvements could be achieved if the final control equipment was optimized. In addition, many facilities audited had one or more of their advanced control applications "turned off" because they were not performing to expectations. Instead of contributing to process optimization they were degrading the performance of other control strategies.

As a result of these studies, the process industries have become aware that the ***entire*** control loop must be optimized to achieve the overall benefits of process optimization and advanced control. In addition, it has been determined that the control valve assembly has the biggest impact on process variability in

[1] M.J. Oglesby, "Achieving benefits using traditional control technologies," Trans Inst MC Vol 18 no 1, 1996
[2] W.L. Bialkowski, "Texas A&M Instrumentation Symposium for the Process Industries, January, 1996

the control loop. If control valves fail to meet high performance requirements they introduce significant process variability that cannot be eliminated with higher level control strategies.

The process industries have also realized that traditional methods of specifying a valve assembly are not adequate to ensure good process loop performance. Such static performance indicators as flow capacity, leakage, materials compatibility, and bench performance data do not deal with the dynamic characteristics of process control loops. More and more control valve users are focusing on dynamic performance parameters (under actual process load conditions) such as dead band, response times, and installed gain as the means to improve process-loop performance. While it is possible to measure many of these dynamic performance parameters in an open-loop situation (these will be examined later), the impact of these parameters really becomes clear when closed-loop performance is measured.

Closed Loop Performance Testing

Over the past six years, Fisher Controls has conducted closed-loop process variability performance tests on a wide variety of control valve assemblies and transmitters. Figure 1 shows that dramatic differences can exist in the dynamic performance capabilities of various control valve assemblies which supposedly are equal products. In these tests, only the control valve assembly was changed (the rest of the loop remained the same).

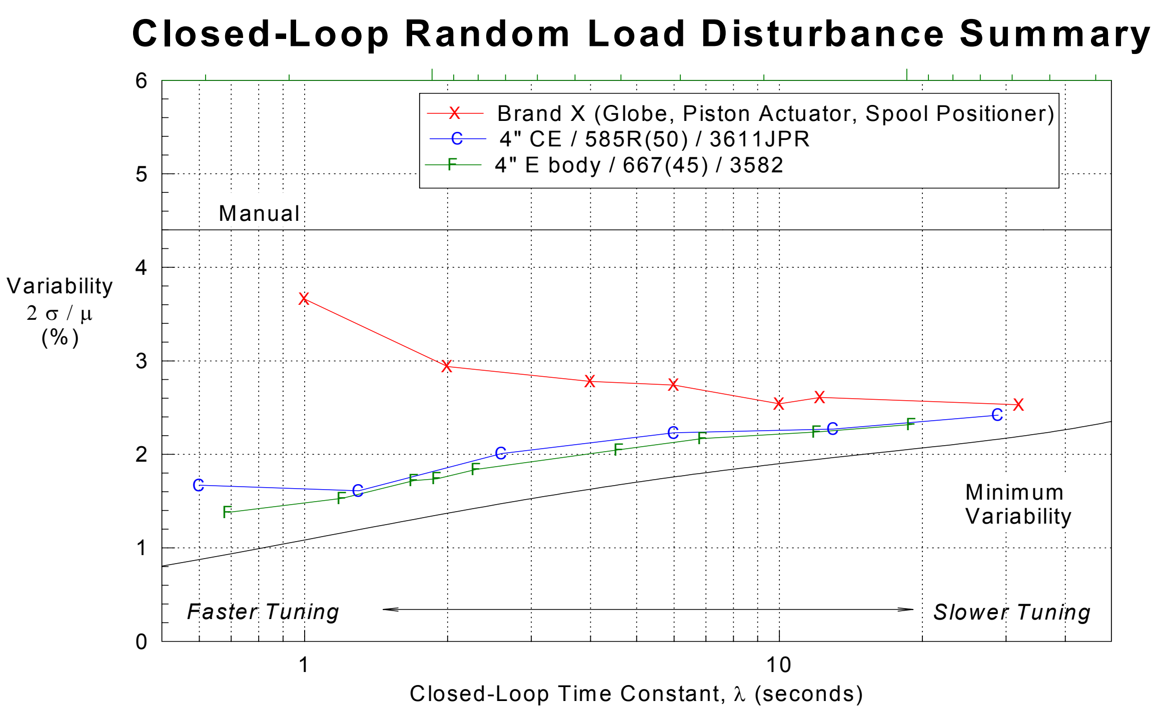

Figure 1. Process Variability Study Results - Control Valve Assemblies

This diagram plots process variability as a percent of the set point variable versus the closed-loop time constant which is a measure of loop tuning. The horizontal line labeled “Manual,” shows how much variability is inherent in the loop when no attempt is made to control it (i.e., open-loop). The line marked “minimum variability” represents the calculated dynamic performance of an ideal valve assembly (i.e., one with no nonlinearities). All real valve assemblies normally fall somewhere between these two conditions.

Figure 1 illustrates that not all valves provide the same dynamic performance despite being considered to be equivalent valves in the same process loop. Note that some valves do a good job of following the minimum variability line over a wide range of controller tunings, which indicates excellent dynamic performance with minimum variability. In contrast, some valves performed less well and actually increase variability as the system is tuned more aggressively for decreasing closed-loop time constants. The variability measured clearly demonstrated that the control valve assembly significantly impacted the overall capability of the process loop.

The financial significance of these test results is enormous. At a closed loop time constant of 3 (figure 1) there is a difference in variability reduction of 1% between the best and worst valve assemblies (the setpoint for these tests was 600 GPM). If the process/material being controlled is worth 25 cents per gallon, the **potential impact on a continuous process would be in excess of $2,100 *per day*!** It is important to note that the valve assemblies tested were new; the differences would be even larger when considering the condition of the valve assemblies typically found in an existing process.

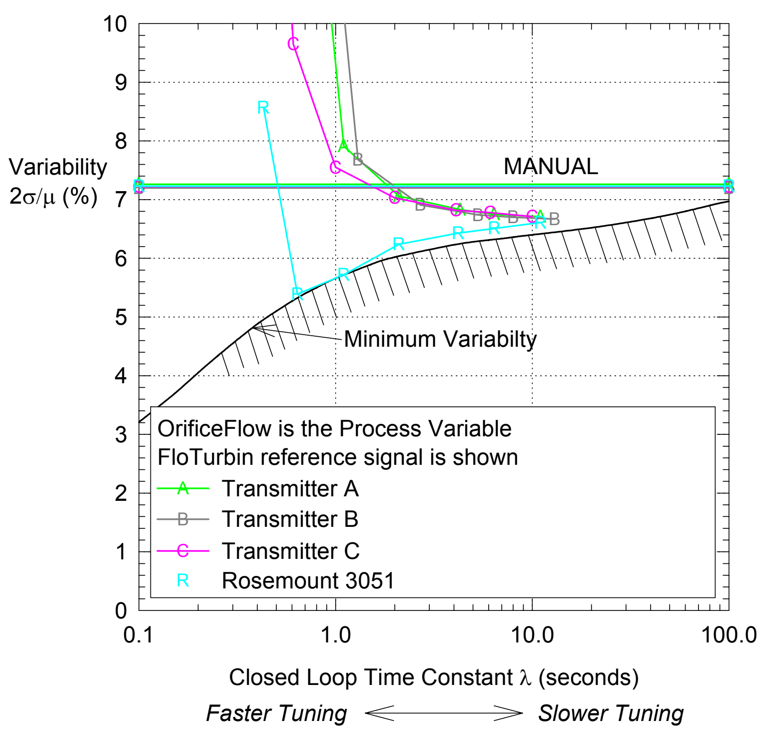

Figure 2. Closed-Loop, Random Load Disturbance Test Summary - Transmitters

In a similar fashion, transmitters tested in closed loop applications show significant differences in dynamic performance. Figure 2 illustrates the results of closed loop testing on various flow transmitters (both analog and “smart”). In this testing, only the transmitter was changed from test to test. The horizontal line labeled “manual,” shows how much variability was inherent in the loop when no attempt is made to control it (i.e., open-loop). The line marked “minimum variability” represents the calculated dynamic performance of an ideal control loop. Again it is clear that transmitter performance is a key consideration for optimal process loop performance. The financial implications of using the “wrong” transmitter are similar to those cited for the control valve.

Real World Process Optimization Examples

Before examining the important performance and design parameters of the final control element, consider the following examples where changes in the control loop equipment had a significant impact on bottom line results:

- A new, higher activity catalyst was charged into a first-stage hydrocracking reactor. After commissioning, reactor temperature variations of +/- 4^{o}F at the catalyst bed were observed with the existing control loop equipment. A loop performance audit revealed that the valve was not capable of responding quickly to process changes. High performance control valve instrumentation was used to replace the instrumentation originally installed on the main hydrogen control valve and the temperature variation was reduced to +/- 0.5^{o}F. As a result, unit production was increased by over 1,000 barrels per day. Based on the average throughput, this optimization was worth $1,400,000 annually. Improvements in the process efficiency were also observed but were not financially quantified.

- In an ammonia plant, poor control of the critical ratio of hot gas and steam resulted in production process upsets and shut-downs. The primary reformer gas flow control valve assembly exhibited slow dynamic response and high friction due to high temperature (697^{0}F) and pressure (678 psi) service conditions. Replacing the control valve assembly and retuning the loop improved control and allowed a 3% increase in the gas flow setpoint. This resulted in an overall production increase of 1% that was valued at $1,200,000 annually.

- In a lube oil blending facility, “off specification” blends (blends requiring blend adjustments) totaled over 11% of the production. A loop performance audit revealed that control loop equipment changes were needed to allow optimal process control. Both valve and transmitter changes were made to six of the loops. Off spec blends have been reduced to 2.3% of the production ….annualized savings of over $7,000,000 are expected.

- In a bleaching process in a pulp and paper mill, a loop performance analysis revealed that the flow variability of the bleach additive (Monox-L) was 8.5% with a TFE lined ball valve. The process dead band of the loop was measured at 10%. Installation of a 0.5% dead band titanium V-notch ball valve assembly reduced the loop flow variability by 3.3%. Monox-L savings due to better control equaled $109,000/year. In addition, paper quality was optimized and environmental concerns were reduced. The additional cost of the titanium valve assembly was recovered in less than two months.

- A chemical reactor was experiencing pressure control problems which were impacting overall control and production. Large fluctuations in process air pressure control resulted in wide swings in the air-to-methanol ratio in the reactor, creating the potential for an explosion. Replacement of the air pressure control valves and system tuning reduced these pressure fluctuations from 15% to 1%. The resulting reduction in process variability increased confidence in the plant’s ability to control the process without explosion. This allowed increased methanol feed rates to the reactor and resulted in production increases of nearly 10%. Financially, this was worth over $750,000 per year.

- The precipitator steam coils in a 1080MW power plant were not functional because of an unstable steam supply control system. This necessitated the use of more expensive "clean" coal. Upcoming deregulation forced the need to use lower cost "dirty" coal; the precipitator system would be required. A loop performance audit indicated that proper valve sizing and selection would yield a much improved installed loop gain. After installation of the equipment, stable automatic operation was achieved for the first time in 20 years! As a result, operating flexibility was improved allowing the use of all varieties of coal. At an average delivered fuel savings of $0.18/MMBTU, an annualized savings potential of $12,000,000 was achieved.

- Performance problems were experienced with the main crude tower feed valve in the preheat train of a crude unit. Pressure fluctuations of greater than +/- 10 psi caused a lower than desired pressure set point to be used in order to prevent lifting the desalter pressure relief valves. Even at the reduced set point, operators eventually resorted to placing the loop in manual in order to prevent the safety valves from being lifted. Tests showed that the original valve had considerable dead band. This valve was replaced with the proper control assembly which reduced the pressure fluctuations to +/- 1 psi. This allowed a higher pressure set point. As a result, production of the unit was increased by 2000 barrels per day for increased revenue of $1,900,000 per year. A further audit of downstream control loops has identified improvements that will result in additional process enhancements.

These examples provide strong evidence that the selection and maintenance of the right final control equipment has a profound economic impact on bottom line performance of the process plants. The ability to perform in the actual process is the real measure of the process equipment. When control loop variability is reduced, the process can be operated closer to the specification or constraint limit. This reduces feedstock and scrap costs, increases productivity (e.g., process throughput and uptime), and improves overall product quality.

Loop Performance Audits

In order to determine the condition of the control loop equipment a disciplined loop performance audit (LPA) should be performed. This is especially important prior to implementation of higher level control strategies or the implementation of process optimization. A good loop performance audit consists of at least three steps as outlined below:

- The first step in a LPA should be a physical review of the installed components compared to the requirements of the P&ID's. In many audits we have found that the installed equipment in the loop was different than that specified. Process data should be reviewed and design basis verified to assure that the right sensor and final control element are being used. A detailed system analysis focused on the installed characteristic and gain of the entire loop can be performed on "profit critical" loops (and problem loops). Any obvious loop equipment problems identified (control valves/instrumentation, transmitters, pumps, piping) should be addressed at this time.

- A another key step of the LPA consists of an on-line assessment of control loop performance. This is performed with a series of controlled, small-step changes in the process. Loop performance is monitored with a special data acquisition system that looks at the process variable, the command signal (input), and the valve travel (optional). The data acquisition equipment used should be capable of at least a 10Hz sample rate. The analysis performed during this step will help identify elements of the loop that need to be examined further in off-line testing.

- If the control valve is identified as a problem during the on-line analysis, a further, more detailed evaluation can be done while the control valve is still installed in the line. For this portion of the LPA, the valve is bypassed (this test can also be done during a shut-down or turnaround) and a series of open-loop tests are performed on the control valve assembly. During this testing a very specific diagnosis of the condition of the valve, actuator, positioner, and I/P is achieved. Similar evaluation of other loop components can also be performed at this time. Again, the data acquisition equipment should be capable of a high sample rate.

With the advent of digital valve controllers and transmitters, the ability to perform continuous, on-line diagnostics of critical field instruments is possible. With these “smart” devices and the appropriate software, not only the health of the equipment, but also the health of the process can be monitored. With these technologies, users can begin to capture the benefits of predictive maintenance practices, which reduce overall maintenance costs while improving process availability and variability. The optimal ***performance*** of the process can be ***sustained*** once the proper loop performance is established with the LPA. A further look at the potential of field-based control devices/systems can be found in reference 5.

Control Valve Performance and Design Parameters

The ability of control valves to reduce process variability depends upon many factors. Research conducted by Fisher Controls has found that the particular design features of the final control element, including the valve, actuator, and positioner, are very important in achieving good process control under dynamic conditions. Most importantly, the control valve *assembly* must be optimized/developed as a unit. Some of the most important design considerations include dead band (friction and backlash), actuator/positioner design, valve style, and valve sizing. A short description of these considerations is provided below. A detailed review of these and other valve design considerations can be found in reference 6.

- ***DEAD BAND (FRICTION AND BACKLASH)***

Dead band is a major contributor to excess process variability, and control valve assemblies are often the primary source of dead band in an instrumentation loop.

Dead band is the general phenomenon where a “band” of controller output values (CO) fails to produce a change in the measured process variable (PV). Figure 3 illustrates a typical dead band situation. When a load disturbance occurs, the process variable (PV) will deviate from the set point. This deviation will initiate a corrective action through the controller and back through the process; however, an initial change in controller output produces no corresponding corrective change in the process variable. Only when the controller output has changed enough to progress through the dead band does a corresponding change in the process variable occur.

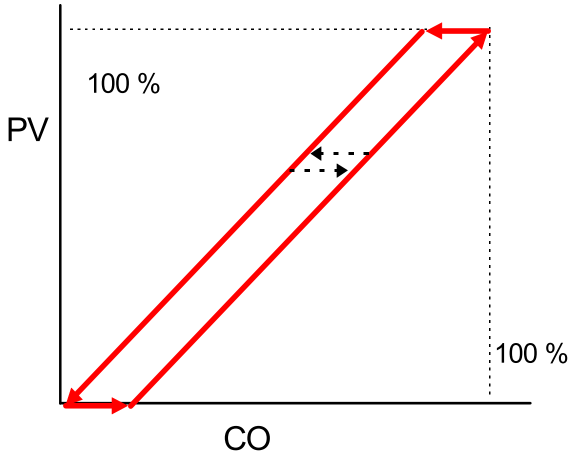

Figure 3: Process Dead Band

Any time the controller output reverses direction, the controller signal must again pass through this dead band before any corrective change in the process variable will occur. The presence of dead band essentially ensures that the process variable deviation from the set point will have to continue to get larger until it is big enough to get through the dead band. Only then can a corrective action occur.

Friction and backlash in the control valve are the two most common contributors to dead band. Since most control actions for regulatory control consist of small changes (e.g., 1% or less), a control valve with excessive dead band may not respond to these small changes. A well engineered valve should be able to respond to signals of less than 1% in order to provide effective reduction in process variability. It is not uncommon, however, for some valves to exhibit dead bands as great as 5% or more.

In a recent performance audit of a gas plant in a refinery, it was found that 30% of the valves had dead bands in excess of 4%. Indeed, over 65% of the loops audited had dead bands greater than 2%. Another audit performed in a gasoline blending unit revealed that over 50% of the loops had dead bands in excess of 2%. In both of these applications the advanced control algorithms were taken out of operation. Many of the control loops had been placed in manual.

As indicated previously, friction must be controlled or accounted for in the control valve design. The overall friction-to-thrust ratio of a control valve assembly is one of the most important considerations in process optimization.

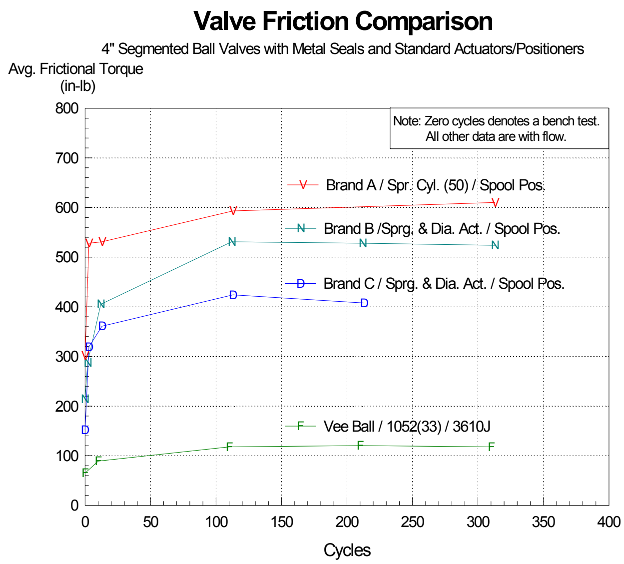

Figure 4: Valve Torque Changes With Use

Figure 4 illustrates the average frictional torque (versus the number of cycles) measured during the testing of four different brands of segmented ball valves. The average frictional torque were measured while stroking the control valves between 5% and 95% open under loaded conditions. While initial frictional torque was relatively low, most of the control valve assemblies exhibited very high frictional torque after a few cycles. As can be seen in figure 4, very high frictional torque (4 to 6 times higher than the best performing product) is a common occurrence.

Figure 5 shows the devastating effect that this higher friction torque can have on the valve's performance. Here the Brand B valve and the Vee Ball® valve from Figure 4 are compared in terms of their response to small step control signal changes to the valve. Note that the Vee Ball® valve assembly is responding to all of the 0.5% input steps for both the valve travel and process variable. On the other hand, the Brand B valve assembly requires input changes of 5% or more to make any significant change in the process variable. Because of the high seal friction, the valve shaft "winds-up" and does not translate motion to the control element. This valve would not be a good choice for process optimization.

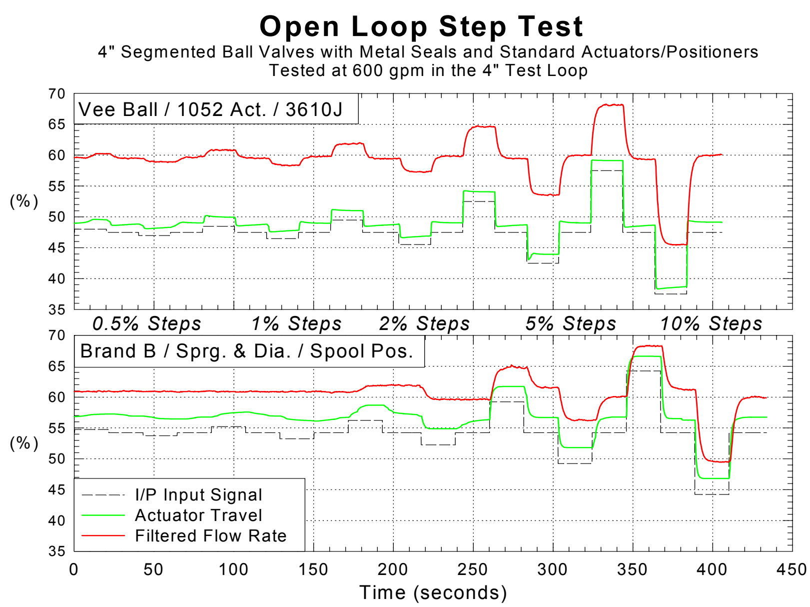

Figure 5. Open Loop Step Test Results

While friction is a difficult phenomenon to eliminate entirely, a well engineered control valve should be able to virtually eliminate dead band due to backlash.

- ***<u>ACTUATOR-POSITIONER DESIGN</u>***

Actuator and positioner design must be considered together; the combination of these two pieces of equipment greatly affect the static performance, as well as the dynamic response of the control valve assembly. More and more, positioners are being purchased for the majority of control valve applications being specified today. Positioners allow for precise positioning accuracy and faster response to process upsets when used with a conventional DCS. With the increasing emphasis upon economic performance of process control, positioners should be considered for every valve application where process optimization (regulatory control) is important.

An important characteristic of a good positioner for process variability reduction is that it has a two-stage design. The first stage (typically a nozzle-flapper) serves as a high-gain, pre-amplifier of signal changes.. This pre-amplification increases the sensitivity of the positioner to very small input signal changes, which is important to achieving reduced process variability.

Single-stage spool valve positioners have become relatively popular in recent times because of their simplicity. Unfortunately, most spool valve positioners omit the first stage pre-amplifier from the design. The input stage of these positioners is simply a transducer module which changes the input signal (electric or pneumatic) into movement of the spool valve. This type of device has less sensitivity to small signal changes which make the one-stage spool valve positioner a poor choice when reduced process variability is important. This poor sensitivity contributes to increased dead band in the control valve assembly.

Attempts are often made to compensate for this poor performance by using spool valves with enlarged ports. This increases the power gain of the device to some extent, but it also dramatically increases the air consumption. Many "high gain" spool valve positioners have static instrument air consumption 5 times greater than high performance 2-stage positioners and still do not achieve equivalent loop performance.

Both lab and field testing have revealed that spring and diaphragm actuators with two-stage positioners outperform piston actuators with single-stage positioners for small (control-size) steps. This is primarily due to a combination of two factors; the lower friction of the spring and diaphragm assembly, and the increased sensitivity and better dynamic performance of the two-stage positioner. This results in shorter dead times and faster response times for the small input changes (2% or less) which are typical for regulatory process control applications (Figure 1 clearly illustrates the impact that positioner/actuator selection has on the reduction of process variability).

Positioners maintain their ability to improve control valve performance for sinusoidal input frequencies up to about one-half of the positioner bandwidth. Because of this, the most successful use of a positioner occurs when the positioner response bandwidth is greater than twice that of the most dominant time lag in the process loop. Some typical examples of where the dynamics of the positioner improve process control are as follows[3]:

- In a distributed control system (DCS) process loop with an electronic transmitter. The DCS and electronic transmitter have time constants that are dominant over the positioner response.

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[3] Gassman, G.W., draft input for the 7th edition of Perry's Handbook, Section 8, Process Control.

- In a process loop with a pneumatic controller and large time constant.

- On springless actuators where the actuator is not usable for throttling control without position feedback.

- In open-loop control applications where best static accuracy is needed.

- When "split ranging" is required to control two or more valves sequentially.

A positioner should not be used in the case where the process controller, the process, and the process transmitter have time constants that are similar or smaller than that of the positioner/actuator.

- ***VALVE TYPE AND CHARACTERIZATION***

The style of valve used and the sizing of the valve have a large impact on the performance of the process control loop. Many people worry about ensuring that the valve size is sufficient to pass the required flow under all possible contingencies; however, it should be noted that a valve which is too large for the specific application is also a detriment to process optimization.

Flow capacity of the valve is also related to the style of valve through the inherent characteristic of the valve style. The inherent characteristic is defined as the relationship between the valve flow capacity and the valve travel when the differential pressure drop across the valve is held constant.

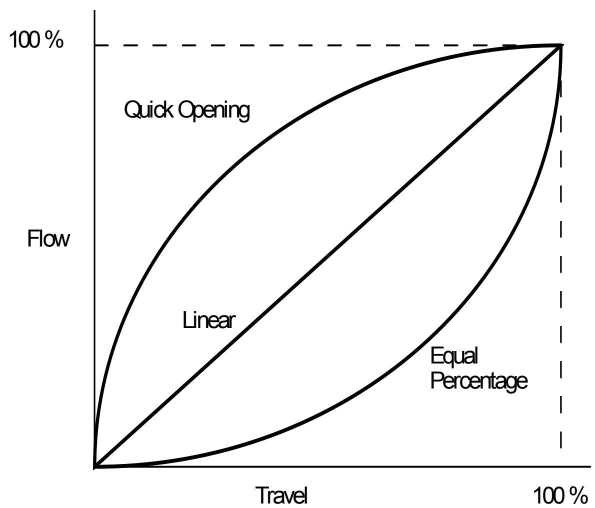

Figure 6: Inherent valve characteristics

Typically these characteristics are plotted on a curve where the horizontal axis is labeled in percent travel while the vertical axis is labeled as percent flow as shown in Figure 6 . Since valve flow is a function of both the valve travel and the pressure drop across the valve, it is traditional procedure to conduct valve characteristic tests at a *constant* pressure drop. This, of course, is not a normal situation in actual practice, but it provides a systematic way of comparing one valve characteristic design to another.

Typical valve characteristics conducted in this manner are named "Linear," "Equal-Percentage," and "Quick Opening" as shown in Figure 6. Under the specific conditions of constant pressure drop, the valve flow becomes only a function of the valve travel and the inherent design of the valve trim. That is why these characteristics are called the "Inherent Flow Characteristic" of the valve. The ratio of the incremental change in valve flow to the corresponding increment of valve travel which caused the flow change is defined as the "Valve Gain.

Knowledge of this inherent characteristic is useful, but the more important characteristic for purposes of process optimization is the ***installed*** flow characteristic of the entire process, including the valve and all the other equipment in the loop. The installed flow characteristic is defined as the relationship between the flow through the valve and the valve travel when the valve is installed in a specific system and the pressure drop across the valve is allowed to change naturally, rather than being held constant. An illustration of such an installed flow characteristic is shown in the upper curve of Figure 7.

The reason for "characterizing" valve gain through various valve trim designs is to provide compensation for other gain changes in the control loop. The end goal, of course, is to attempt to maintain loop gain which is reasonably uniform over the entire operating range; i.e., to maintain a relatively linear ***installed*** flow characteristic.

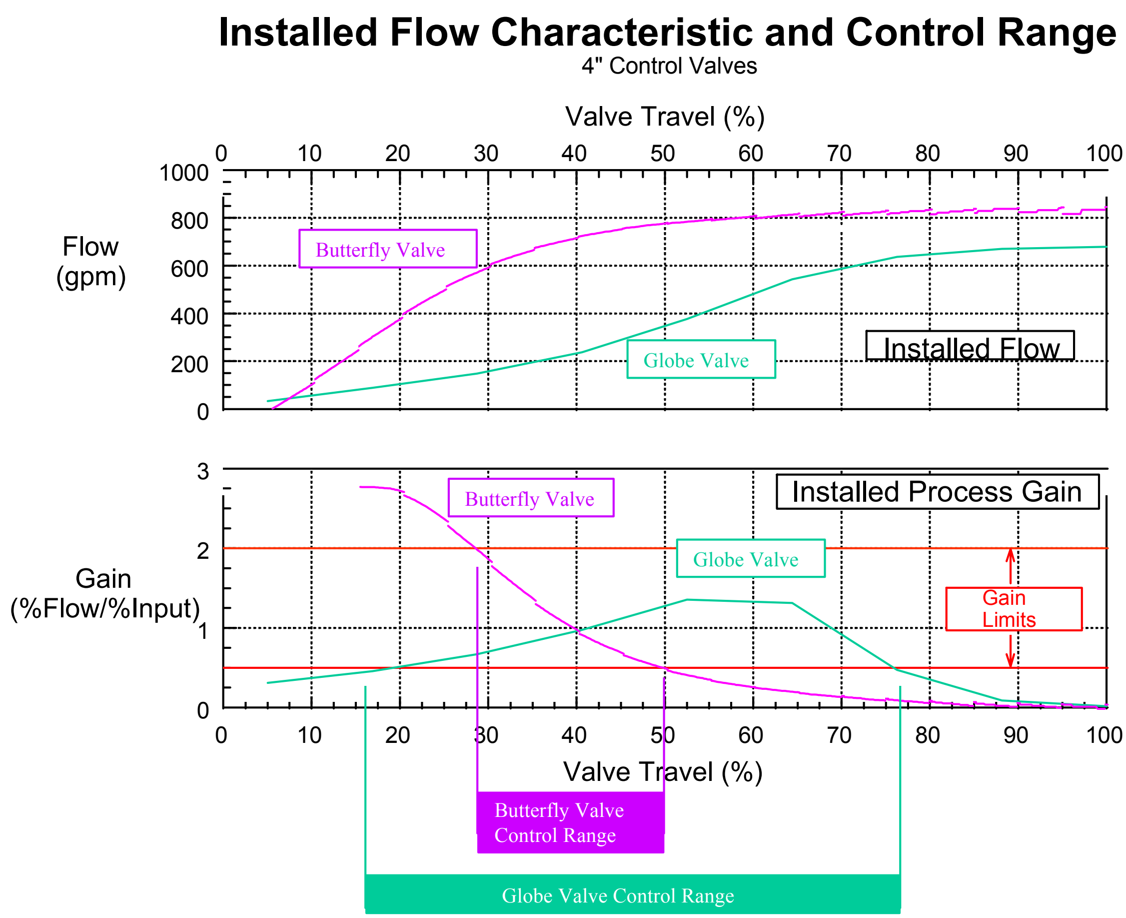

Figure 7: Installed flow characteristic and installed loop process gain.

Theoretically, a loop is tuned for optimum performance at some set point flow condition. As the flow varies about that set point, it is desirable to keep the loop gain as constant as possible in order to maintain optimum performance. If the loop gain change due to the inherent valve characteristic does not <u>exactly</u> compensate for the changing gain of the unit being controlled, then there will be a variation in the loop gain and process optimization will become more difficult. There is also the danger that the loop gain might change enough to cause limit cycling, or other dynamic difficulties.

A common guideline is that the installed loop gain should not vary over more than about a 4:1 ratio, otherwise the dynamic performance of the loop will suffer unacceptably. There is nothing magic about this specific ratio, it is simply one which many control practitioners agree will produce an acceptable range of gain margins in most process control loops. Because of its common acceptance, this guideline forms the basis for the following EnTech gain limit specification[4]:

Loop Process Gain = 1.0	(% of transmitter span)/(% controller output)
Nominal Range: 0.5 - 2.0 %/%	(Note 4:1 ratio)

This definition of the "loop process" includes everything in the loop except the controller; i.e., the product of the gains of the control valve, the unit being controlled, and the transmitter. Since the valve is part of the "loop process", it is important to select a valve style and size which will produce an installed flow characteristic which is sufficiently linear to stay within the specified gain limits over the operating range of the system. If too much gain variation occurs in the control valve itself, it leaves less flexibility in adjusting the controller. Good practice is to keep as much of the loop gain in the controller as possible.

The installed characteristic curves in Figure 7 were obtained in an operating control loop. With the controller in manual (open-loop), flow is established with the load valve held constant at a nominal operating condition. The test valve input signal is then varied throughout its normal travel range and the corresponding flow is measured. The flow versus valve travel results of this test are shown in the upper portion of Figure 7. This curve is known as the "installed flow characteristic." The slope of this curve can be determined at each travel increment. This slope of the installed flow characteristic curve is the "installed process gain." The installed gain is shown as the curve in the lower plot of Figure 7.

The butterfly valve shown in Figure 7, when installed in this loop, can only operate inside the acceptable gain specification limits over an approximate travel range of about 20% (*control range*). In contrast, a globe valve in this same loop has a *control range* of over 55% (with proper adjustment of the 4:1 range, the globe valve can actually have a control range of 70%).

In the valve performance hierarchy, globe valves demonstrate the widest *control range* and V-notch ball valves have the next widest *control range*. Eccentric plug type valves generally have a narrow *control range* when compared to globe and V-notch ball valves. Butterfly valves typically have the narrowest *control range* and are generally suited for only fixed load applications. Most importantly, process optimization requires that a valve style and size be selected that will stay within the selected gain limit range over the widest possible operating range. Many times, selection of an inappropriate valve style results in poor dynamic loop performance.

4 "Control Valve Dynamic Specification" (Version 2.1, 3/94), EnTech Control Inc., Toronto, Ontario

- ***VALVE SIZING***

A typical problem encountered when trying to optimize process performance is the general practice of oversizing valves. This typically results from the common practice of using line size valves, especially with high-capacity rotary valves, as well as the conservative addition of multiple safety factors at different stages in the process and facility design. Often plants are designed for some “future” capacity and the control loop equipment is sized for the future situation. This practice is detrimental to process optimization and results in performance losses that far exceed the cost of the proper loop equipment.

Oversizing the valve hurts process variability in several ways. First of all, the oversized valve puts too much gain in the valve, leaving less flexibility in adjusting the controller. Best performance always results when most of the loop gain comes from the controller. The second way that oversized valves hurt process variability is that an oversized valve is likely to operate more frequently at lower valve openings where seal friction may be greater, particularly in rotary valves.

In addition, regardless of its actual inherent valve characteristic, a severely oversized valve tends to act more like a quick-opening valve with very high installed valve gain in the lower lift regions. This often causes the valve to operate outside the acceptable process gain limits which results in a very narrow control range for the valve.

Finally, since an oversized valve produces a disproportionately large flow change for a given increment of valve travel, this phenomenon can greatly exaggerate the process variability associated with dead band due to friction, etc.

Summary

Process optimization means optimizing the *entire* process, not just the control algorithms utilized in the control room equipment. The equipment in the process loop must be capable of performing to realize the benefits of advanced control. It makes no sense to install an elaborate process control strategy and hardware instrumentation system capable of achieving 0.5% or better process control and then implement that control strategy with a 5.0% control valve. Audits performed on thousands of process control loops have provided strong proof that the final control element plays a significant role in achieving true process optimization.

Control valves are a highly engineered product and should not be treated simply as a commodity. While traditional valve specifications certainly play an important role, it is crucial that valve specifications also address real dynamic performance characteristics if true process optimization is to be achieved. The performance of the control loop in the process should be the prime consideration when specifying equipment.

It is important to realize that process optimization begins and ends with optimization of the *entire* loop. A formal loop performance audit should be used to identify and correct problems associated with loop hardware. You cannot simply treat separate parts of the loop individually and expect to achieve coordinated loop performance. Likewise, you cannot hope to evaluate the performance of any part of the loop in isolation. Isolated tests under non-loaded, bench-type conditions will never provide the kind of performance information that is obtained from testing the hardware in actual process conditions.

With the advent of digital valve controllers and transmitters, the ability to perform continuous, on-line diagnostics of critical field instruments is possible. With these “smart” devices and the appropriate software, not only the health of the equipment, but also the health of the process can be monitored. With these technologies, users can begin to capture the benefits of predictive maintenance practices, which reduce overall maintenance costs while improving process availability and variability. The optimal ***performance*** of the process can be ***sustained*** once the proper loop performance is established with the loop performance audit.

While additional effort and consideration is required to address all of these issues, the payoff in optimized service and performance will have a major impact on profitability as was demonstrated by the examples provided in this paper.

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