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IMPROVEMENTS TO A CENTRIFUGAL COMPRESSOR SURGE CONTROL SYSTEM

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ABSTRACT

A surge control system for a natural gas centrifugal compressor station has been modified in order to reduce shutdowns caused by high discharge temperature and provide a more robust and stable operation. The process consists of a compressor driven by a 14 MW gas turbine and recycle piping, a 16" recycle valve, a PLC based surge control algorithm, a flow measurement element, and a compressor differential pressure transmitter. The control objective is to manipulate the recycle valve to maintain flow through the compressor to a setpoint determined from the differential pressure across the compressor. Field tests were conducted to measure the open loop process dynamics of the valve, piping, compressor and transmitters. From the test data, the relevant process dynamics were determined enabling the development of a first order plus dead time model of the system. The process dynamics are complex due to the gas dynamic effects of the station piping and tend to exhibit inverse and time delayed behavior. Large variations in process gain also create problems with obtaining a consistent flow response under different operating conditions. A stability analysis was completed and the control system was redesigned with several enhancements including derivative control, flow signal filtering, process linearization, and improved controller programming techniques. The results of the modifications are the compressor does not shut down when subjected to transients from other units, the compressor can be started against high head conditions, and the closed loop response time is ten times faster than the previous system. The new system has been in operation since May 1997.

NOMENCLATURE

b	bias of surge setpoint
BCL	secondary or backup control line
c	surge controller output characterization constant
D	derivative control action
E	control error
FOPDT	first order plus dead time model
I	integral control action
K_c	proportional gain
K_p	process gain
m	slope of the surge setpoint
P	proportional control action
P_d	differential pressure developed by the compressor
Q	suction to impeller eye differential pressure at compressor
R	flow control setpoint, also referred to as the Surge Control Line (SCL)
SLL	surge limit line
t	time
t_d	process dead time
U	control output
τ_I	integral time constant
τ_d	derivative time constant
τ_p	process time constant

INTRODUCTION

The surge control system is essential to the safe and reliable operation of a centrifugal compressor. The system must provide protection to the compressor by preventing the flow from decreasing to the point where surge occurs, a phenomena that can cause mechanical damage to the internals of the compressor. The system must also provide a stable operation during transient events such as startup, shutdown, and speed changes initiated by requirements to control pressure on the pipeline.

This paper describes a new surge controller design that has been implemented on a compressor currently operating on the NOVA Gas Transmission pipeline. The compressor was subject to frequent shutdowns on high discharge gas temperature caused by excessive recycling. The purpose of the design project was to implement a design that reduced excessive recycling and improved the reliability of the unit without sacrificing the level of surge protection provided by the existing control system.

The design process was conducted in four steps:

1. Identification of process dynamics through field testing and development of process model

2. Formulation of realistic performance objectives.

3. Design and offline testing of a control system to accomplish those objectives using the process model from step 1.

4. Field implementation and testing.

The entire system is represented schematically in Fig 1. A suction to eye differential pressure transmitter produces a signal proportional to flow through the compressor. The flow signal (Q) is monitored by a programmable logic controller (FIC). The controller also monitors the differential pressure across the compressor and calculates a flow setpoint (R) from this signal. The controller produces an output signal (U) that drives the recycle valve to the desired position. The changing recycle valve position modulates the flow of gas to the compressor.

The block diagram representation of the process control system is shown in Fig 2. For the purpose of process identification and control system design, the process consists of the control valve, compressor, check valve, piping, and flow measurement combined into one block labelled Gp. The control is labelled Gc.

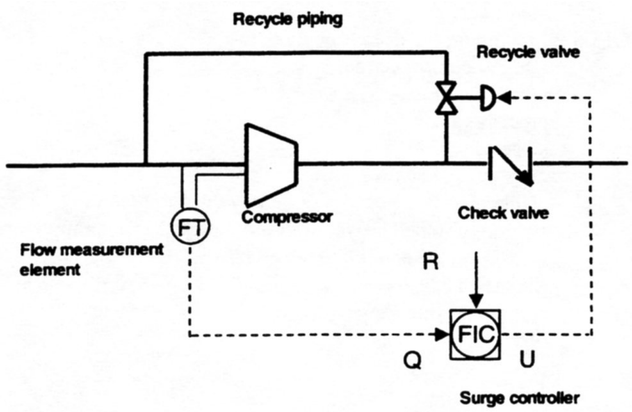

Fig. 1 Process and instrumentation schematic

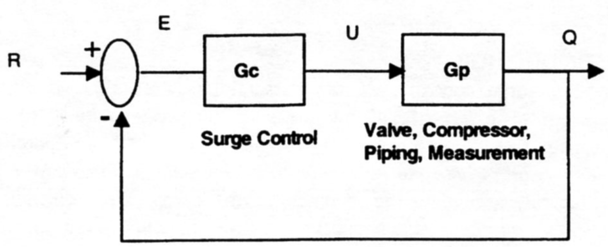

Fig. 2 Process control block diagram

PROCESS IDENTIFICATION

The process dynamics were identified using both open and closed loop tests.

For the open loop tests, the surge controller was placed in manual, the compressor speed was held constant, and the valve position was changed incrementally by 10%. During each change the valve position and measured flow were recorded at 100 ms intervals. The results of the response tests during opening and closing of the valve are shown in Fig. 3 and Fig. 4, respectively. The time of initiation of the valve movement has been adjusted to correspond with 300 ms for each of the tests in order to enable the comparison of time constants between the various test cases.

A first order plus dead time model (FOPDT) was applied to each open loop response test. The FOPDT method is a standard modelling approach used in the process control field to provide a simple model for process control studies. This model is described by the following first order linear differential equation:

$$\tau_p \frac{dQ}{dt} + Q = k_p \cdot (u(t - td)) \tag{1}$$

The process gain k_p, dead time t_d, and first order time constant τ_p were determined for each test. Figure 5 shows the first order plus dead time approximation plotted against flow for one of the tests.

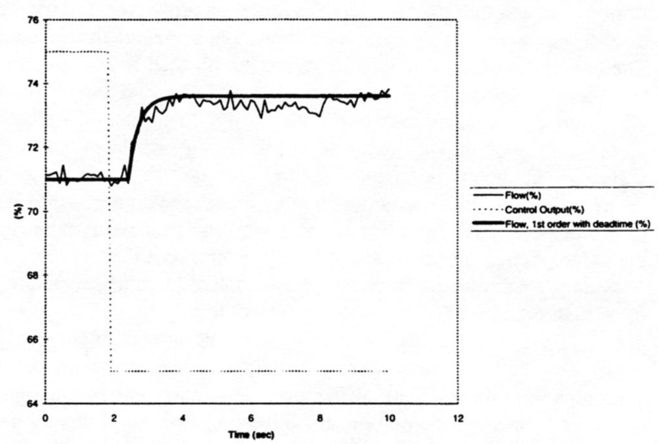

Fig. 5 First order plus dead time approximation

The process parameters were identified using the following methodology:

k_p: change in flow / change in control output

τ_p: time for flow to change from starting value to 63% of final value

t_d: time from control output change to initial flow change.

In Fig. 6, the process gain for each test is plotted against valve position, illustrating the nonlinear character of the process to changes in valve position.

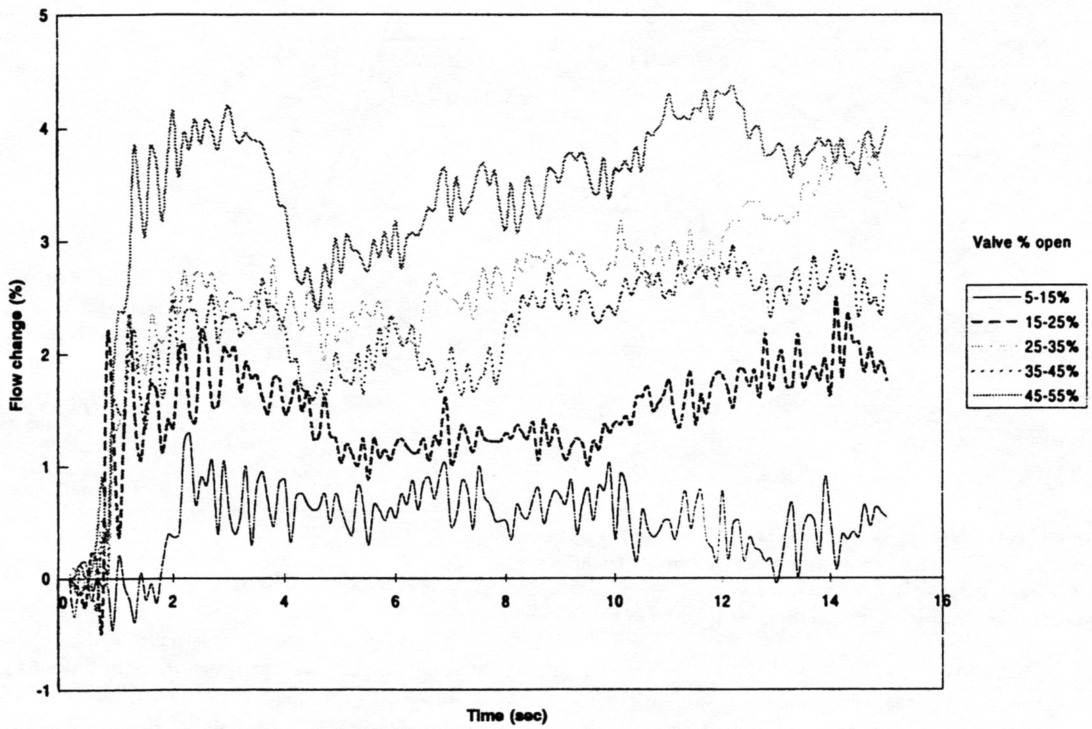

Fig. 3 Open loop test results during valve open command

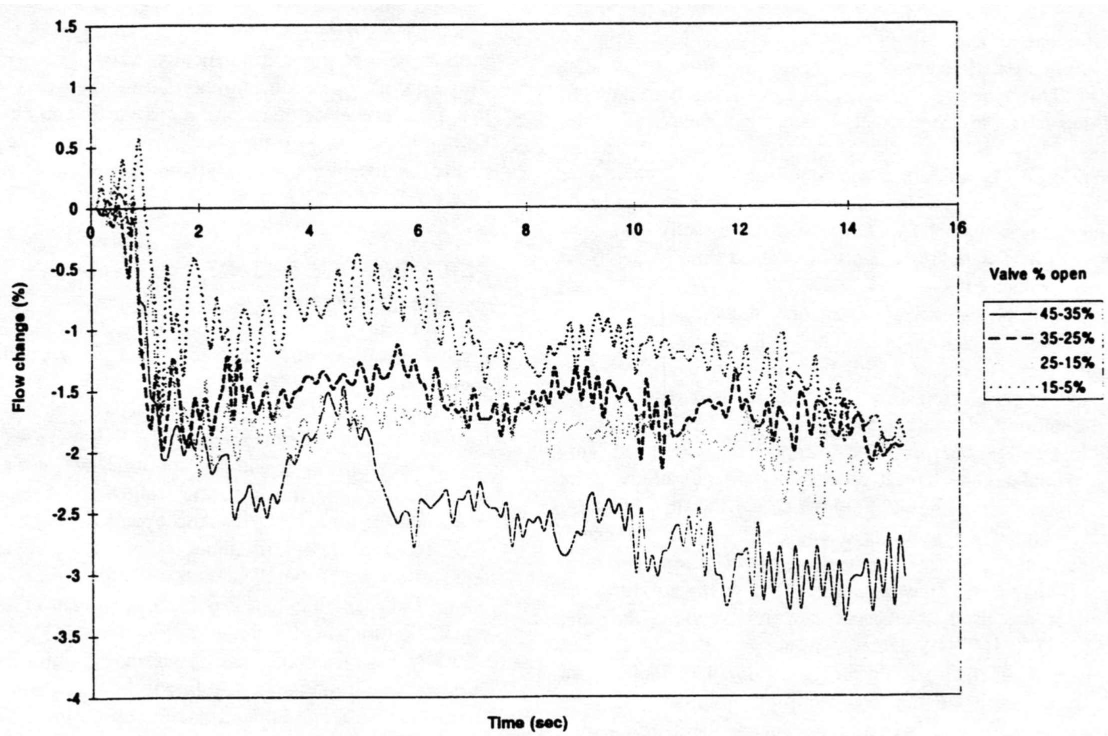

Fig. 4 Open loop test results during valve close command

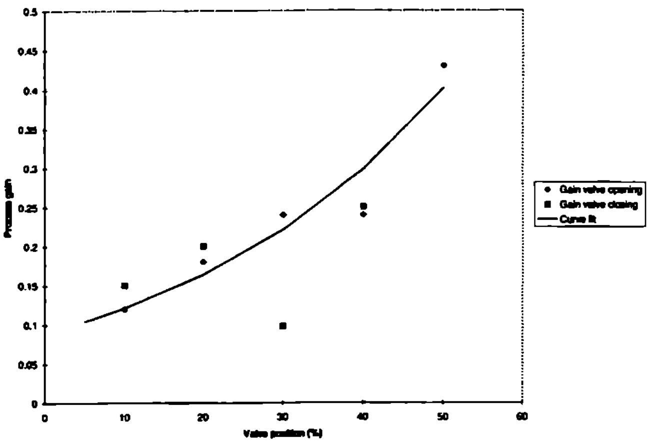

Fig. 6 Process gain identification

As the valve opens the process gain increases, from 0.1 to 0.4 at valve openings of 0 % to 50%, an effect primarily a result of the equal percentage installed flow characteristic of the valve.

The total process time constant and dead time were found to be:

$\tau_p = 0.3$ s

$t_d = 0.8$ s

The 0.8 second dead time can be divided into the following effects:

Gas dynamics	0.07
PLC input/output	0.10
Valve hysteresis	0.63

The gas dynamic dead time contribution was estimated by dividing the piping length from the valve to the flow transmitter sensing points (20 m) by the speed of sound in gas, roughly 300 m/s to get 0.07 s of delay. The PLC input/output delay was measured in the control processor.

The remainder of t_d was attributed to friction in the valve that prevents the valve from moving immediately following a change in air pressure on the valve actuator. This value was not directly measured but was calculated by subtracting the previous dead time values from the total observed value of 0.8 s.

The response tests also reveal an inverse response that occurs about 5 seconds after the control output change is initiated. When the valve is opened, the flow first increases and then decreases five seconds later before recovering again. This effect is a result of gas dynamics in the station and mainline piping.

A closed loop test was performed by ramping the unit speed. This test effectively simulates the largest process disturbance likely to be encountered during actual operation. Figure 7 shows the flow response resulting from reducing the unit speed enough to cause the check valve to close, at which point oscillations in flow occurred. The ratio between the amplitude of the flow oscillations versus the amplitude of the control output was used to measure the process gain while the check valve was closed. This value was found to be about 1.4 and represents a four times increase in process gain. The increase in process gain is a result of the smaller pipe volume since the downstream piping is isolated from the system by the check valve. Open loop tests could not be conducted in this mode without causing a unit shutdown since the unit cannot be operated with the check valve closed for more than 30 seconds due to rising discharge temperature.

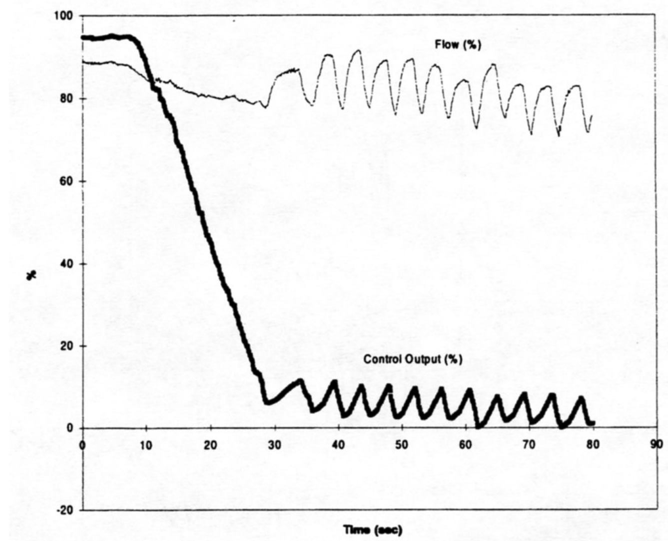

Fig. 7 Closed loop test results during speed ramp

A final closed loop test was performed by changing the setpoint. The results of this test are shown later in order to compare as found performance to improved performance. The time required for the as found system to return flow to the setpoint is about 30 seconds. The flow changes in a highly overdamped manner, suggesting the possibility of increasing the controller gain in order to reduce the response time.

An important result of the open and closed loop tests is that the process gain changes dramatically when the check valve is closed. This information becomes important for designing a control algorithm that is robust enough to provide a sufficiently fast response under normal operation with the check valve open while maintaining stable control at the higher process gain that occurs with the check valve closed.

PERFORMANCE OBJECTIVES

The general objectives of the surge controller are to prevent the unit from surging and to prevent the unit from shutting down on high discharge temperature due to excessive recycling during station transients.

These general performance objectives are related to the response time and stability of the system in the following ways:

A very slow over-damped controller will not respond fast enough to transients and will sometimes allow the flow to decrease to the point that the emergency backup open loop control must be initiated. This can then result in unnecessary recycling and high discharge temperature shutdown. The as found condition of this unit was over-damped to such an extent that the backup control line (BCL) would be reached during small unit speed reductions.

A very fast acting controller may exhibit excessive overshoot causing the surge control trip line to be reached the same as above or it may result in un-damped oscillations, especially during operation with the check valve closed, again leading to excessive recycling.

In order to provide a response that is satisfactory under both of the above scenarios, the following measurable performance characteristics were chosen both from general control system tuning experience and from knowledge of the process characteristics.

1. Set the closed loop response characteristic to slightly over-damped during normal operation when the check valve is open.

2. Allow the closed loop response to be under-damped with a decay ratio of no more than 1/8 during transient operation (starting and stopping) when the check valve is closed.

These easily measurable objectives can be utilized during the design phase of the new controller and may be modified during field testing.

CONTROLLER DESIGN

For the purposes of determining the controller design, a simple mathematical model was derived from the process identification results. The model was used to simulate the process flow responding to a change in recycle control valve position. The purpose of the model was not to detail every aspect of the complex nature of compressor/pipeline dynamics, but rather to provide a means of determining controller gains within a +/- 10% range of final field commissioning values and to test alternative controller designs such as derivative gain. The simple FOPDT model very successfully modelled the actual installation to the extent that the controller could be coarsely tuned in the office rather than in the field. This delivers a large benefit when field tuning as the initial tuning values will provide stable operation while final tuning values are determined.

The general arrangement of the surge control system is based on the system for variable speed compressors developed by M.H. White, 1972, and further discussed by G.K. McMillan, 1983. The surge controller uses flow from the flow transmitter as the measured variable and differential pressure across the compressor from the differential pressure transmitter as the controller setpoint. This arrangement appears as a flow control loop with a calculated setpoint. The setpoint is described by the following equation:

$$R = m \cdot P_d + b \qquad (2)$$

This control is implemented through a traditional PI closed loop feedback algorithm. Operating at or near the SCL requires that the surge controller be tuned for a very fast response. However, a compromise must be reached between fast response and stability of the control loop. A very rapid decrease in flow, for example when compressor speed is reduced quickly, can result in excursions to the left of the SCL and in some cases could cause compressor surge. The conventional method of incorporating a backup or secondary open loop override control algorithm is used to address these fast load disturbances. When the operating point reaches the BCL, the open loop control algorithm immediately strokes the recycle open to move the operating point to the safe region right of the SCL. While it is an integral portion of the surge control system, the open loop response is upsetting to the process as it causes the system to be inefficient by recycling gas. It also increases the likelihood that the recycled gas will be overheated and cause a high discharge gas temperature shutdown.

For these reasons, optimizing the closed loop response of the controller is of prime importance in providing stable, efficient and reliable operation. Figure 8 shows the relationship between the SLL, SCL and BCL.

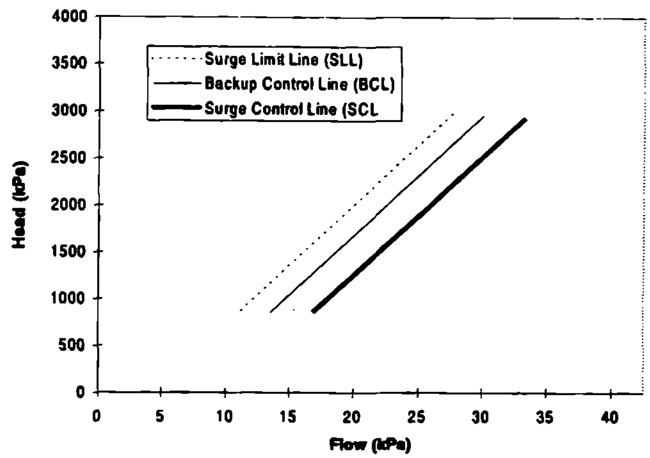

Fig. 8 Surge control lines

While the general design of the controller is based on the work done by M.H. White and others, a number of enhancements were made to the key concepts of the design. The objective of the enhancements was to better the trade-off between performance and stability, in effect improving the robustness of the control. That is to say, improve the responsiveness of the control while retaining a high degree of stability.

To address sluggish control response the normal course of action would be to increase the controller gain (proportional response) and increase the speed of the integral (or reset) action. Although these changes increase responsiveness of the controller, both actions are moves in a de-stabilizing direction. The challenge then becomes one of finding a way to increase the controller proportional and integral action while retaining stability. Normally this can be accomplished by adding derivative action, supplying a stabilizing effect provided that it is not used in excess. Derivative action is not commonly used in flow control loops because of the high degree of flow measurement noise. Noise is a disturbance that is not true (i.e. electrical noise) or is too fast to be reacted to. Excessive noise will be amplified by the derivative action causing the controller to oscillate, sometimes uncontrollably.

During the initial design phase signal noise can be minimized by using proper piping design principles upstream and downstream of the flow measurement device. However, often noisy flow measurement signals are a fact of life and they usually result in low controller gains and poor performing surge control systems. Controller performance can be increased if noise could be decreased to acceptable levels. Measurement noise can be decreased if signal filtering were added. Filtering of the signal does not come without a price - it adds a phase lag to the loop.

The key to the success of this tuning approach is to combine derivative control with a flow filter time constant that is just large enough to stabilize the signal without adding excessive lag to the loop. The phase lag caused by the signal filter is compensated for by the predictive nature or phase lead added through derivative action. To further enhance the control, the flow signal input to the PLC incorporates a signal splitter that allows independent amounts of signal filtering for the closed loop algorithm and the open loop algorithm. The open loop mode can therefore have a smaller time constant on its filter with a resulting shorter time lag and in this way still react very quickly to rapid load disturbances while at or near the SCL. Further controller aspects are discussed in more detail as they pertain to the test unit:

Signal Filtering

The flow transmitter has an adjustable damping feature that should be set to minimum, resulting in a response time of 50 to 60 ms. The surge control execution time is 100 ms.

The flow signal filters were programmed in the PLC using a first order backward difference equation. The time constants for the filters were set at:

closed loop: 300 ms
open loop: 100 ms

Figure 9 shows the effect of a 300 ms filter on the input signal.

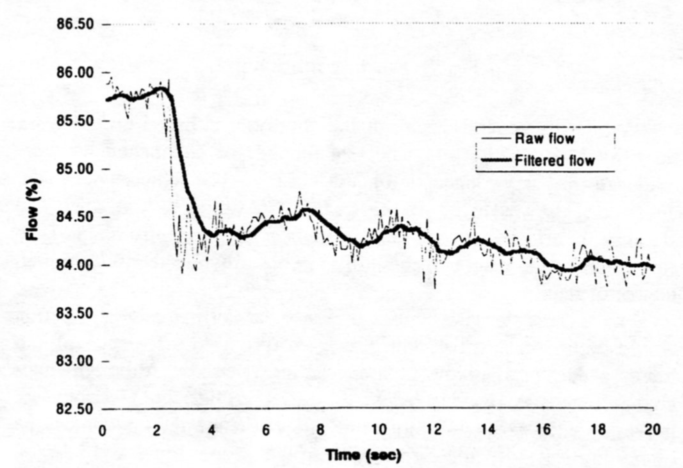

Fig. 9 Flow measurement filtering

PID Controller Settings

The PLC uses the following equation to calculate an output from controller error:

$$U = k_c \cdot \left(E + \frac{1}{\tau_i} \int_0^t E \cdot dt + \tau_d \cdot \frac{dE}{dt} \right) \qquad (3)$$

Data from the test unit indicated that the existing system was very sluggish, primarily because of low integral action and a noisy flow measurement signal. The existing settings were K_c=0.7, τ_i = 1.8 s, τ_d = 0 s.

Figure 10 displays a simulation of performance of the surge controller using derivative action, with and without flow filtering. The controller settings were adjusted to:

K_c = 0.8
τ_i = 0.5 s
τ_d = 0.2 s

Performance in this situation is enhanced when filtering is used. In this case, derivative control is not recommended without process variable filtering.

The simulated closed loop response to a setpoint change was created by numerical integration of the combined control and process equations. The characteristics of the process were:

τ_p = 0.3 s
t_d = 0.8 s
k_p = 1.4

The oscillations that occur in the "no flow filter" case are a result of the excessive gain of the derivative term combined with the loss of phase lead resulting from the dead time in the process. The filter limits the bandwidth of the PID controller effectively reducing the gain of the controller derivative term at higher frequencies. This results in the elimination of the oscillations in the diagram below.

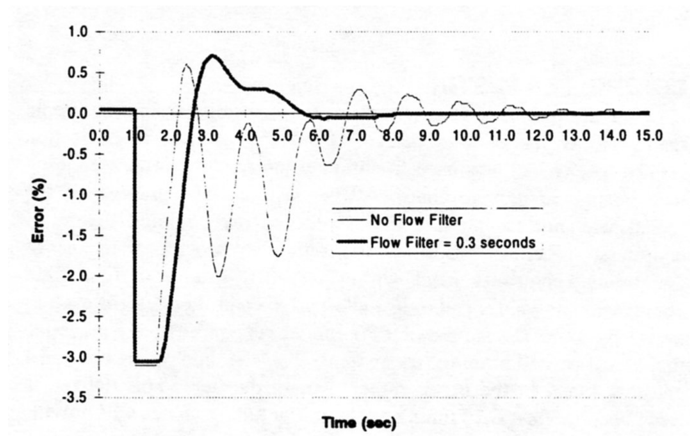

Fig. 10 Effect of flow filtering and derivative control

The final tuning settings implemented in the field were:

K_c = 0.8
τ_i = 0.5 s
τ_d = 0.2 s

Process Linearization

The surge controller can be more accurately tuned if the flow response from the process is relatively linear throughout the entire control valve range. On the test unit, as is the case with many of the NGTL installations, the installed characteristic of the recycle system (valve and piping) is non-linear. Most commonly the recycle valve trim is equal percentage resulting in the same characteristic in the system as a whole. To address the changes in process gain as the valve strokes open, a software characterizer function was added to the control system. This feature tends to produce a more linear overall system response. The feature can be enabled or disabled as required and is adjustable to suit the application. Figure 11 shows the characterizer effect.

The following equation describes the characterization function:

$$V = c^U \qquad (4)$$

where c is a constant chosen by the user to determine the degree of characterization required.

Open Loop Response

As discussed above, while the open loop response is an essential portion of the overall surge control system, it can have unsettling effects on the process. Throwing the recycle valve wide open and then slowly ramping the valve closed, can result in high gas temperature shutdowns and unnecessary recycling of gas. The open loop response was modified to only open the recycle valve the amount necessary to move the operating point to the safe region and then allowing the improved closed loop response to effectively control the recycle valve.

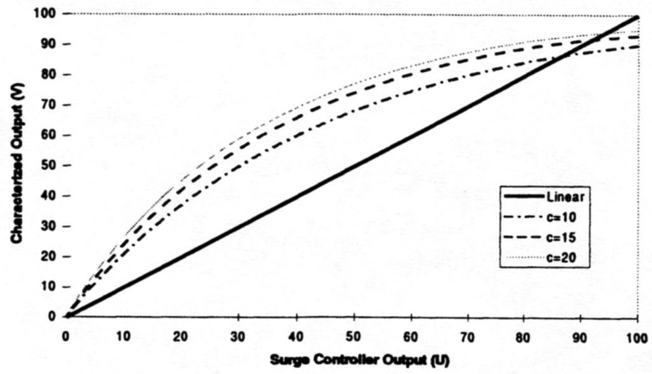

Fig. 11 Control valve characterization

RESULTS

Steady State Error

The new system restricted the loop error to less than 1% in steady state conditions. This represented a 3 to 1 ratio improvement on the existing system.

Response To Setpoint Step Change

The following performance criteria were chosen in order to compare controller responses after a step change in the setpoint was introduced:

1. Integral of absolute error multiplied by time (ITAE)
2. Rise time to initially reach setpoint after a step change
3. Overshoot - as a % of the initial step change
4. Settling time - time for absolute error to be less than 1% after a step change

The measurements in Table 1 were taken after a setpoint step change was introduced on both the existing system and the new system.

Table 1
Performance Comparison: As Found to New System

	As found	New	Improvement ratio
ITAE	3779	326	11 : 1
Rise time (s)	11.5	3.2	4 : 1
Overshoot (%)	75.0	10.6	7 : 1
Settling time (s)	96	4	24 : 1

Figures 12 and 13 are graphical descriptions of comparative performance.

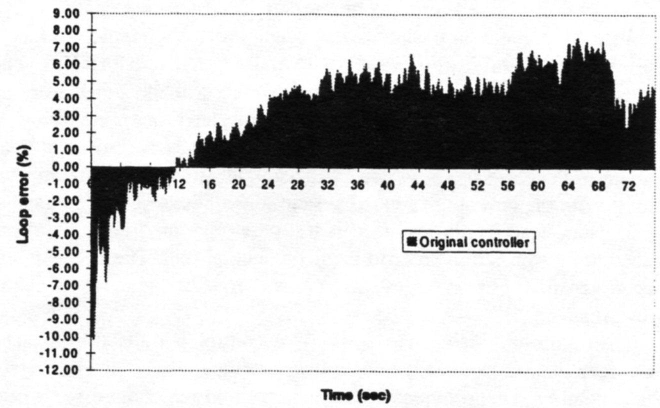

Fig. 12 Existing surge controller - cumulative loop error

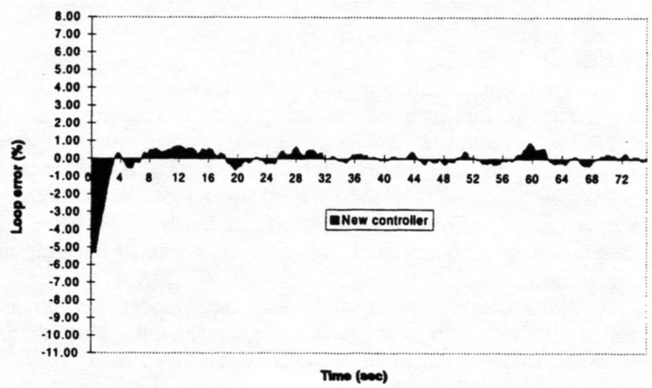

Fig. 13 New surge controller - cumulative loop error

Response To Load Disturbance

A large load disturbance is placed on a unit when that unit's speed is reduced while the speed on parallel units remains constant. This will result in a large loss in flow to the unit decreasing in speed while the other unit will increase flow.

The new surge control system performed better than the existing system in keeping the operating point at or to the right of the SCL. However, ultimately there are load disturbances that are large enough or fast enough that the closed loop controller will not be able to keep the operating point from crossing the SCL. Under these conditions, the open loop control will respond by opening the recycle valve and moving the operating point to the right of the SCL.

Control Effort

The minimization of error places a premium on controlling to the setpoint. As a consequence the new control system is more responsive and sensitive to control loop error. Observers of the new system will find that the controller output will fluctuate more, even in steady state conditions, when compared to the existing system.

Pipeline System Impacts

A robust control system is capable of providing acceptable response over a wide range of process conditions. In the pipeline application the range of process conditions include pressure disturbances, unit discharge check valve position, start up and run down conditions and remote compressor speed setpoint changes. The ability to prevent unit shutdowns when process conditions change provides a significant benefit to overall system performance and operation. The new system with derivative control and signal filtering is highly stable and responsive under all expected process conditions.

With the previous system, other parallel units had to be slowed down to decrease the head across the station prior to starting another unit. With the new surge control system, the unit can be started against maximum head conditions. In this test case, pipeline linepack did not need to be adjusted in order to bring on another unit. The unit is much less sensitive to changes in process conditions and pipeline disturbances.

In addition, stable and responsive control permits the distance between the SCL and SLL to be reduced without compromising safety or reliability. This provides additional room on the compressor wheel map to be available, allowing the system operation to be more flexible.

CONCLUSIONS

The following conclusions are drawn from this investigation:

1. The performance of a centrifugal compressor can be significantly improved by implementing inexpensive software modifications to the surge control system. These performance achievements result in reduced high discharge temperature shutdowns, improvements in efficiency through reduced levels of recycling, and the ability to immediately start additional units without lowering the station head.

2. A simple first order plus dead time model can provide qualitative value in assessing surge control performance characteristics prior to field implementation. This includes

a) use of flow filtering

b) effect of derivative gain

c) verification of tuning settings

3. PID control with flow signal filtering can provide a more robust system than PI control. Higher levels of dampening are achieved with the derivative control when the process gain increases, such as when the unit operates with the discharge check valve closed.

ACKNOWLEDGEMENTS

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