Optimizing Turbomachinery Controls Symposium

Houston, TX
June 2013
CCC Global Facilities

Des Moines

Abu Dhabi

Amsterdam

Italy

China

Australia

Brazil

Moscow
CCC collaborates with customers to deliver reliable solutions for critical turbo machinery control applications that result in tangible economic benefits.
Participant Introductions

Any specific objectives for this session?
Agenda – Day 1

• Optimizing Turbomachinery Control / Financial impact
• Compressor Classifications
• Turbocompressor Control Techniques
  - Measuring Distance from Surge
  - Auxiliary Equipment Selection
  - Specialized Antisurge Control Response
  - Performance (Process) Control Integration
  - Stand-alone Compressor Simulation
  - Using Recycle for Process Control
• Working with Turbocompressor Networks
  - Parallel Compressor Loadsharing
    • Base-loading, Equal Flow, Equidistant Techniques
• TrainView Web Workshop – Attendee Operation
Agenda – Day 2

• Steam Turbine Control
  - Turbine Start-up & Shutdown Automation
  - Speed & Extraction Control Techniques and Demo

• Application Examples
  - LNG Liquefaction - Refrigeration Compressors
  - NGL Fractionation Facilities & Bayu Undan Example
  - Ammonia-Urea Unit Applications

• Process Control vs. Safety Shutdown Systems
  - System Availability and Fault Tolerance
  - API Standards Update

• Tips for Specification Writing
Controls

Machinery

Turbomachinery Controls

Antisurge
Performance
Speed
Extraction
Loadsharing
Quench
Flare

Centrifugal Compressors
Axial Compressors
Steam Turbines
Gas Turbines
Hot Gas Expanders
Cryogenic Expanders
Power Recovery Trains

LNG
Offshore
FPSO
NGL
GTL
Gas Transmission

PTA
Olefins
Refining
Ammonia
Methanol
Air Separation

Process
Major Challenges in System Design

- Accurately Representing Compressor Operation:
  - Variable process conditions
  - Variable gas composition
  - Lack of flow elements in some sections

- Control Element Selection & Sizing
  - Valve selection & sizing
  - Transmitter range specification
  - Recycle piping arrangement
  - Designing for recycle cooling/quench

- High Speed Response:
  - Flat performance curves (especially near surge)
  - Desire for operating stability & efficiency
  - Desire to maximize the operating envelope
Major Challenges in System Design

• Dealing with Control Loop Interaction
  - Antisurge to antisurge decoupling
  - Antisurge to performance control decoupling
  - Coordinated mode switching between machines

• Loadsharing for Compressor Networks
  - Parallel, Series, & Compound arrangements

• Integrating Multiple Limiting Variables
  - Compressor limits
  - Driver limits
  - Process limits

• Maximizing Overall System Availability
  - Providing reliable control system redundancy
  - Effectively dealing with field equipment faults
Financial Impact of Optimizing Turbomachinery Control Systems
Lifecycle Costs

30-year life cycle costs for a 20,000 HP compressor

* Costs in constant dollars

Source: “Experiences in Analysis and Monitoring Compressor Performance”
Ben Duggan & Steve Locke
E.I. du Pont, Old Hickory, Tennessee
24th Turbomachinery Symposium
Lifecycle Costs

30-year costs per a 1,000 HP

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Ben Duggan & Steve Locke
E.I. du Pont, Old Hickory, Tennessee
Presented at the 24th Annual Turbomachinery Symposium.
Machinery Controls - Retrofit Economics

- While turbo-compressors & turbines can last > 30 years, control systems can be obsolete < 10 years
- Newer control systems can offer:
  - Better performance & integration capabilities
  - Improved machinery protection
  - Better system availability
    - More reliable digital electronic components
    - Improved system redundancy schemes
    - Improved self-diagnostic capabilities
    - Integrated predictive maintenance tools
- ROI can be extremely attractive due to the potential for:
  - Increases in process throughput
  - Increased “insurance” against catastrophic failures
  - Energy savings
Compressor Types & Classifications
Rotary Screw Air Compressors

- Also Positive Displacement Machines
- Normally 200HP to 1000HP
Centrifugal Compressors

- Widespread use, many applications
- Gas is accelerated outwards by rotating impeller
- Can be built for operation as low as 5 psi, or operation as high as 10,000 psi (35 kPa - 55,000 kPa)
- Sizes range from 200 HP to 95,000 HP
Cross Section of Horizontal Split

- Compressor inlet nozzle
- Thrust bearing
- Journal bearing
- Shaft and labyrinth seal
- Impeller inlet labyrinth seals
- Casing (horizontally split flange)
- Impellers
- Drive coupling
- Discharge volutes
- Compressor discharge nozzle
- Compressor inlet nozzle
Multi-Section with Bull Gear
Compressor volute
Pinion shafts
2nd Stage impeller
Drive coupling
Labyrinth seals
Main drive gear
Inlet guide vanes
Compressor volute
Pinion shafts
2nd Stage impeller
Drive coupling
Labyrinth seals
Main drive gear
Inlet guide vanes
Compressor volute
Barrel-Type Centrifugal
Barrel-Type Centrifugal
Axial Compressors

- Gas flows in direction of rotating shaft
- Can be built for lower pressures only
  - (10 to 100 psi discharge pressures)
- Very efficient for high flow applications
- Not as common as centrifugals
Compressor System Classifications

Single-Section, Three-Stage

Single-Case, Two-Section, Six-Stage

Parallel Network

Two-Case, Two-Section, Six-Stage

Series Network
Compressor Control Solutions
Critical Success Factors

1. Reliability
   - Minimize effect and duration of process disturbances

2. Efficient Operation
   - Prevent surge, overspeed, overheating, and associated damage
   - Automate startup and shutdown
Critical Success Factors

1. Reliability
   - Maximize process throughput
   - Minimize setpoint deviation

2. Efficient Operation
   - Optimize loadsharing
   - Minimize antisurge recycle or blow-off
   - Operate at lowest energy level
Critical Success Factors

• Prevent unnecessary process trips and downtime

1. Reliability

• Minimize effect and duration of process disturbances

2. Efficient Operation

• Prevent surge, overspeed, overheating, and associated damage
• Automate startup and shutdown
<table>
<thead>
<tr>
<th>Service</th>
<th>Risk – Cost of Outage PER EVENT</th>
<th>Major Parts Cost and Production/sales/environmental impact</th>
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<tbody>
<tr>
<td></td>
<td>Estimated Repair time</td>
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<tr>
<td>Lean Gas Re-compressor</td>
<td>1 week</td>
<td>$2-3 MM USD</td>
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<td>No LPG production</td>
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<tr>
<td>C3 Refrigeration Compressor</td>
<td>1-2 weeks (Compressor)</td>
<td>$4-6 MM USD (Comp)</td>
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<tr>
<td>(Drier Pre-cooler)</td>
<td>7-10 days (Steam Turb)</td>
<td>&gt; $6 MM USD (ST)</td>
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<tr>
<td>Off-Gas Compressor</td>
<td>1 week (Compressor)</td>
<td>$2-3 MMUSD (Comp)</td>
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<td></td>
<td>7-10 days (Steam Turb)</td>
<td>$3 MM USD (ST)</td>
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<td></td>
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<td>LNG production stops</td>
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<td>Fuel Gas Compressor</td>
<td>2 weeks (Compressor)</td>
<td>$6-9 MM USD (Comp)</td>
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<td></td>
<td>7-10 days (Steam Turb)</td>
<td>$3 MM USD (ST)</td>
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<td>Frame 9Es must use FFF and not meet NOx requirements set by SCE</td>
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<tr>
<td>BOG Booster Compressors</td>
<td>1 week</td>
<td>1.5-2MM USD</td>
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<tr>
<td>Regeneration or Excess Gas</td>
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<tr>
<td>Compressor</td>
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**Critical Success Factors**

1. **Reliability**
   - Maximize process throughput
   - Minimize setpoint deviation

2. **Efficient Operation**
   - Optimize loadsharing
   - Minimize antisurge recycle or blow-off
   - Operate at lowest energy level
Increased Production

Ethylene Plant Example / Charge-Gas Compressor

- Every 0.1 bar (1.5 psi) increase in 3\textsuperscript{rd} stage Pd provides annual production increase of approx. 1.5%.

- An average reduction of 0.1 bar (1.5 psi) in 1\textsuperscript{st} Stage Ps (coil outlet pressure) provides an estimated 1% increase in Yield.


**Increased Production**

- **Steel Mill Example**
  - Hot Metal production is directly proportional to the mass flow of “wind to the furnace”
  - Loss of flow during stove-filling procedures in the range of .5 to 1.5% of total flow (…production)
Developing the Compressor Curve

- Pd: Discharge Pressure (P2)
- ΔPc: Differential Pressure (Pd - Ps) or (P2 - P1)
- Rc: Pressure Ratio (Pd/Ps) or (P2/P1)
- Hp: Polytropic Head

Compressor curve for a specific speed N1
Integrating Multiple Limiting Variables

Your turbomachinery control system must move smoothly in and out of compressor and driver limits while providing effective process control and stability.
What is Surge?
Surge Description

- Flow reverses in 20 to 50 milliseconds
- Complete surge cycle in 300ms to 3 seconds
- Compressor vibrates
- Temperature rises
- “Whooshing” noise
Consequences of Surge

- Rapid flow and pressure oscillations
- Exponential temperature increase
- Process instability & lower efficiency
- Potential machine damage
- Trips may occur
- Conventional instruments and operators may fail to recognize surge
Consequences of Surge

Damage on axial

← compressor – stator

Damage on axial

compressor - rotor →
Developing the Surge Cycle

- As pressure builds, resistance goes up, operating point “draws” the perf curve
- Compressor reaches surge point A
- At this point the potential energy in the vessel exceeds the capability of the compressor
- Because $P_v > P_{d(max)}$, the flow reverses
- Pressure drops and forward flow is reestablished

$P_d = \text{Compressor discharge pressure}$

$P_v = \text{Vessel pressure}$

- From A to B…..20 - 50 ms…….. Drop into surge
- From C to D…..20 - 120 ms…… Jump out of surge
- A-B-C-D-A…..0.3 - 3 seconds…… Surge cycle
Factors Leading to Surge

- Machine/Process startup and shutdown
- Operation at reduced throughput
- Operation at heavy throughput with:
  - Process upsets / Load changes
  - Operator errors
  - Gas composition changes
  - Cooler problems
  - Filter or strainer problems
  - Driver problems / Power loss / Trips
- Surge is not limited to times of reduced throughput, it can and does occur under full operation
Calculating Distance from Surge
Fact:
- The more inaccuracy in the measurement of distance to surge, the farther we’ll need to stay from the surge limit to ensure reliable operation.

Challenge:
- The compressor surge limit moves with compressor inlet conditions: Temperature, Pressure, and gas composition.
- The antisurge system should provide a distance to surge calculation that is invariant of any change in inlet conditions.
- This will lead more reliable surge protection while reducing the required surge control margin:
  - Larger turndown range on the compressor
  - Reduced energy consumption during low load conditions.
Typical compressor maps include: \((Q_s, H_p), (Q_s, R_c),\) or \((Q_s, P_d)\) coordinates, where:

- \(Q_s\) = Suction flow / can be expressed as actual or standardized volumetric flow
- \(H_p\) = Polytropic Head
- \(R_c\) = Compressor Ratio \((p_d / p_s)\)
- \(p_d\) = Discharge pressure of the compressor
- \(p_s\) = Suction pressure of the compressor
- \(k_s\) = Exponent for isentropic compression

These maps are defined for (1) specific set of inlet conditions: \(P_s, T_s, MW & k_s\)

Assume isentropic process (no changes in gas composition or in system entropy)
Limitations of OEM Coordinate Systems

- These coordinates are **NOT** invariant to suction conditions as shown

- For control purposes we want the SLL to be presented by a single curve for a fixed geometry compressor
Distance From Surge is a Moving Target

- Most all process compressors will also have to operate in off-design conditions.
- Automatic antisurge control must be able to operate under varying operating conditions:
  - Suction Temperature
  - Suction Pressure
  - Gas Composition
  - Rotational Speed

Design Nitrogen Off-gas

<table>
<thead>
<tr>
<th>Design</th>
<th>Nitrogen</th>
<th>Off-gas</th>
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<tbody>
<tr>
<td>MW</td>
<td>MW</td>
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<tr>
<td>$P_s$</td>
<td>$P_s$</td>
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<tr>
<td>$T_s$</td>
<td>$T_s$</td>
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<tr>
<td>$k_s$</td>
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Distance From Surge is a Moving Target

• A map based on polytropic head & volumetric flow coordinates is much closer to invariant when inlet conditions change, however:
  - There are no $H_p$ transmitters available
  - Gas analyzers are generally slow & unreliable for use in continuous control
  - Multiple analyzing points would be necessary to calculate $Z_s$ and $Z_{avg}$

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Equations for Polytropic Head & Volumetric Flow:

\[ H_p = Z_{avg} R \; T_s \times \frac{R_c^\sigma - 1}{\sigma} \]

\[ Q_s^2 = Z_s R \; T_s \times \frac{\Delta P_{os}}{P_s} \]

Where:

- \( Z_{s,d} \) = Compressibility in Suction, Discharge
- \( Z_{avg} \) = Average Compressibility = \( \frac{Z_s + Z_d}{2} \)
- \( T_s \) = Suction Temperature
- \( R_c \) = Compression Ratio = \( \frac{P_d}{P_s} \)
- \( P_d \) = Discharge Pressure
- \( P_s \) = Suction Pressure
- \( R = \frac{R_u}{MW} \) = Gas Constant
- \( MW \) = Molecular Weight
- \( R_u \) = Universal Gas Constant
- \( \sigma = \frac{k - 1}{k \eta_p} \) = Specific Heat Ratio = \( \frac{C_p}{C_v} \)
- \( k = \) Specific Heat Ratio
- \( C_p \) = Specific Heat at Constant Pressure
- \( C_v \) = Specific Heat at Constant Volume
- \( \eta_p \) = Polytropic Efficiency
- \( \Delta P_{os} \) = Differential Pressure across orifice plate at Suction
Reduce each equation by a factor $A$:

$$A = Z_s R T_s$$

$$\frac{H_p}{A} = \frac{Z_{avg} R T_s \cdot \frac{R_c^\sigma - 1}{\sigma}}{Z_s R T_s} = h_{p, red}$$

$$\frac{Q_s^2}{A} = \frac{Z_s R T_s \cdot \frac{\Delta P_{os}}{P_s}}{Z_s R T_s} = q_{s, red}^2$$

From experience, we know that the ratio of $Z_{avg}/Z_s$ varies negligibly. Assuming it to be constant over the compressor operating range:

$$h_{p, red} = \frac{R_c^\sigma - 1}{\sigma}$$

$$q_{s, red}^2 = \frac{\Delta P_{os}}{P_s}$$
σ Calculation

The relationship between pressure & temperature in polytropic compression theory leads to a practical way to measure \( \sigma \):

\[
\frac{T_d}{T_s} = \left( \frac{P_d}{P_s} \right)^\sigma
\]

Solving for \( \sigma \):

\[
\frac{\log \left( \frac{T_d}{T_s} \right)}{\log \left( \frac{P_d}{P_s} \right)}
\]

Calculating \( \sigma \) improves system accuracy when:
- Gas composition varies
- Compressor efficiency changes (air compressors)
Reduced coordinates define a performance map which:

- Is invariant to changing inlet conditions
- Has one surge limit point for a given rotational speed and compressor geometry
- Permits calculation of the operating point without obtaining molecular weight and compressibility measurements

\[ h_{p,\text{red}} = \frac{R_c^s - 1}{S} \]

\[ q_{s,\text{red}}^2 = \frac{\Delta P_{\text{os}}}{P_s} \]

Slope_{\text{OPL}} = \frac{h_{p,\text{red}}}{q_{s,\text{red}}^2}
The controller continuously calculates:

The ratio of the slope of the Surge Limit Line & the slope of the Operating Point line:

\[ S_s = \frac{\text{Slope}_{\text{OPL}}}{\text{Slope}_{\text{SLL}}} \]

The distance between the operating point and the surge limit:

\[ d = 1 - S_s \]

Variable \( S_s \) defines the angular distance from the operating point to the surge point
The controller calculates deviation from the SCL as:

\[ \text{DEV} = d - b_1 \]

Where:

\[ d = 1 - S_S \]

\[ b_1 = \text{the relative distance between the SLL & the SCL} \]

\text{DEV} represents the distance between the operating point and the surge control line.
The surge parameter is defined as:

\[ S_s = \frac{f_1(h_r)}{q_{r_{op}}} \]

Non-linearity in the Surge Limit Line can be accommodated using a function based on a piecewise characterization of either map axis.

The function \( f_1 \) returns the value of \( q_{r{SLL}} \) on the SLL for input \( h_r \).
Advantage of Invariant Coordinates / Illustration (1)

NOT invariant coordinates \((H_p, Q_s)\)

Invariant coordinates \((h_r, q_r^2)\)

Where:
- \(H_p\) = Polytropic head
- \(Q_s\) = Volumetric suction flow
- \(Q_r^2\) = Reduced flow squared
- \(h_r\) = Reduced head

Choose the right coordinates for each antisurge application
Hydrogen Recycle Compressor

**Invariant Coordinates**

**Variant Coordinates**

- **Curve 1:** MW = 4.62; Ps = 6.033 kg/cm² a
- **Curve 2:** MW = 5.90; Ps = 6.800 kg/cm² a
- **Curve 3:** MW = 7.90; Ps = 14.900 kg/cm² a
- **Curve 4:** MW = 8.20; Ps = 6.800 kg/cm² a
- **Curve 5:** MW = 9.70; Ps = 14.900 kg/cm² a
- **Curve 6:** MW = 10.8; Ps = 14.900 kg/cm² a
Surge Testing
Surge Testing: a Daily Routine

“High Speed Recording of Inlet Flow”
Why Surge Test?

• Maximizing operating envelope
  - Validating predicted compressor maps
  - Establishing accurate surge limit (or safe operating point near surge)

• Field-testing compressor responses in process conditions through field devices
  - Actual transmitter field calibrations
  - Actual flow element installation
  - Actual process gas (in some cases)

• Determining signature of surge for proper configuration of surge detection system
Why Surge Test?

• Testing surge control system response in difficult process conditions
  - Ensuring system is capable of preventing surge and breaking surge cycle (if necessary)

• Field performance data available for future benchmarking
  - Can the machine develop enough head?
  - Can the compressor operate over the complete range needed?

• Preventing unexpected surprises after commissioning
Surge Detection

Pressure and Flow Variations During a Typical Surge Cycle

• Surge signature should be recorded during commissioning
• Rates of change for flow and pressure transmitters should be calculated
• Thresholds should be configured slightly more conservatively than the actual rates of change during surge
• Surge is detected when the actual rates of change exceed the configured thresholds
• The following methods have been used:
  - Rapid drop in flow
  - Rapid drop in pressure
  - Rapid increase in temperature
  - Rapid drop in flow & pressure
**Typical CCC Surge Test**

- Procedure typically takes place with the compressor “off-process” or in nearly full recycle.
- CCC engineer starts with predicted surge limit & default surge control margin.
- Antisurge valve held in manual in partially open position.

![Diagram showing Initial State, Predicted Surge Limit Line, Default Surge Control Line, and Operating Point.](image-url)
Typical CCC Surge Test

- The surge limit line (SLL) is moved to the right in a safe distance from the predicted surge limit.
- The antisurge control margin \((b_1)\) is then set to "0" (no margin).
- The antisurge valve is slowly closed, moving the operating point to the temporary SCL/SLL (see next slide).
Typical CCC Surge Test

- Antisurge controller is placed in AUTO
- CCC engineer slowly decreases “K” thereby increasing the slope of the SCL/SLL
- The operating point follows the SCL/SLL moving ever closer to the actual surge limit
- This action is repeated in small steps, letting the process stabilize between movements.

Graph showing:
- "Safe" Surge Limit Line
- Temporary Surge Control Line
- Operating Point

Equation: $b_1 = 0$
Typical CCC Surge Test

- At the actual surge point, a sudden drop in flow will occur
- Recycle Trip™ and Safety On™ work to break the surge cycle and reestablish stable control in a safe operating region
- Established K value (slope of the SLL) is evaluated in relation to existing data
Typical CCC Surge Test

- Test data is adjusted as necessary and entered into the antisurge control parameter set.
- CCC engineer then enters a new surge control margin ($b_1$) appropriate for ongoing operation of the controller.
- Safety On™ is then reset, moving the SLL back to the newly established and now field tested position.
• In this example, (2) “identical” compressors were surge tested during CCC system commissioning
• The predicted surge limit on Train A turned out to be significantly to the left of the predicted curve
• Thus, the test identified the potential for a wider operating range to be established on Train A
Is Surge Testing Required?

• Adequate set-up of the antisurge system without surge testing during commissioning is achievable in most cases, however:

  - Our experience indicates that there are often significant differences between predicted compressor performance documentation (maps) and field test data collected

• Field service engineers should perform testing only with coordination & cooperation of the complete commissioning team
Factors Affecting Surge Control Margin
The Approach to Surge is Fast

- Performance curves are almost invariably flat near surge
- Even small changes in compression ratio cause large changes in flow
- Suction or discharge pressure controller will accelerate the operating point in the direction of surge
Factors Affecting Surge Control Margin

Larger Surge Control Margin
- Inability to surge test
- Less than ideal piping configuration
- High noise-to-signal ratio on system inputs
- Compromised recycle valve response
- Possible large process disturbances
- Flat curves

Smaller Control Margin
- Surge testing with varying speeds and process gas conditions
- Ideal piping configuration
- Clean and repeatable transmitter signals
- Fast & accurate recycle valve response
- No large process disturbances
- Steep performance curves
A leading engineering firm was contracted to evaluate the effect of controller execution rates on system response:

- A dynamic controller-in-the-loop simulation was built
- An antisurge control application was selected
- Digital controllers were compared to an analog controller
- Analog controller tuned for minimum overshoot
- Digital controllers got the same tuning parameters
- All trials use the same disturbance: closure of discharge block valve
Analog vs. Digital Controller at 2 Executions per Second

- Compressor surged
- Large process upset would have resulted

Analog controller

Digital controller (2 executions per second)

Tuning same as analog controller
Analog vs. Digital Controller at 10 Executions per Second

- Compressor almost surged
- Control system would have to be set up with bigger surge control margin

Tuning same as analog controller
Analog vs. Digital Controller at 25 Executions per Second

- Response of CCC digital controller nearly identical to the analog controller
- Adding specialized response algorithms to PI control will allow even smaller surge margins

![Graphs showing response of analog and digital controllers](image-url)
Piping Layout Considerations
Piping Layout Considerations

- Piping layout significantly influences controllability
- The primary objective of the antisurge controller is to protect the compressor against surge
- This is achieved by lowering the resistance to compressor flow through opening the antisurge valve
- Dead-time & lag-time should be minimized
- This is achieved by minimizing the volume between three flanges:
  - Discharge flange of the compressor
  - Recycle valve flange
  - Check valve flange
Using a Single Antisurge Valve

- In order to protect section 1 the antisurge valve needs to be opened
- The volume between compressor discharge, check valve and antisurge valve determines the dead time and lag time in the system
- Large volume significantly decreases the effectiveness of the antisurge protection

Results:
- Poor surge protection
- Large surge margins
- Energy waste
- Process trips because of surge
The piping configuration for section 2 is excellent for surge protection.
- Minimum volume between the three flanges.
- The piping configuration for section 1 is not ideal as there is a large volume of process gas to be de-pressurized.

**Results:**
- Poor surge protection
- Large surge margins
- Potential energy waste
- Potential process trips due to surge
Improving Control Response

- Compressor 1 has ideal piping for surge protection as the volume between the three flanges is minimized.
- The piping layout for Compressor 2 is common but not optimum from an antisurge system response point of view.
  - The cooler creates additional volume and decreases the effectiveness of the antisurge control system.
Optimum Surge Protection

- This compressor has ideal piping configuration for surge protection as the process volume is minimized between the three flanges for all sections.
Which Piping Configuration Would you Choose?

- Configuration #1 has minimum volume between the flanges and is the best layout for antisurge control reaction time.
- Configuration #2 requires one less cooler, less capital investment is lower and cooler maintenance costs are reduced.
- Configuration #2 will require bigger surge control margins.
Specialized Control Response
Coping with the High Speed of Approaching Surge

- Maximize speed of response
  - Transmitters
  - Valves
  - System volumes
  - Controllers

- Utilize specialized control responses
  - “Predictive” response algorithms
  - Automated open loop (Recycle Trip™)
  - Adaptive surge control line
  - Control loop decoupling
Basic Antisurge Control System

- The antisurge controller UIC-1 protects the compressor against surge by opening the recycle valve.
- Opening of the recycle valve lowers the resistance felt by the compressor, reducing $R_c$ and increasing flow.

This takes the compressor away from surge.
The Surge Control Line (SCL)

- When the operating point crosses the SCL, PI control will open the antisurge valve.

- PI control will give adequate protection for small disturbances.

- PI control is tuned to provide stable control during steady state operation on the SCL.
The Recycle Trip® Line (RTL)

• Disturbance arrives - the operating point moves past the SCL initiating Proportional & Integral (PI) control action

• If the operating point hits the RTL, the conclusion is:
  - We are close to surge
  - The PI controller was too slow to catch the disturbance
  - Move the valve now!

• An open loop response is triggered to open the antisurge valve at maximum actuator speed
The Recycle Trip® Line (RTL)

- For large, short duration disturbances a fixed ramp out of the open loop response may hold the antisurge valve open too long
  - Could result in an unnecessary loss of process flow
  - Antisurge controller should utilize its distance from surge calculation to accelerate the exponential decay

- After the operating point moves back to the safe side of RTL, the open-loop function should be ramped out
  - PI controller will stabilize the operating point on the SCL until normal process flow is re-established
The Recycle Trip® Line (RTL)

Output to Valve

Total Response

PI Control Response

Recycle Trip® Response

Time

- Total response of the controller is the sum of the PI control and the Recycle Trip® action

Benefits:

- Wider operating envelope is created due to reduced SCL margin requirement
- Energy savings due to smaller required surge margins
- Surge can be prevented for virtually any disturbance
• Open loop response is the fastest way to get the antisurge valve open
  - But, open loop control lacks the accuracy needed to precisely position the antisurge valve

• By definition, open loop corrections of a fixed magnitude are often either too big or too small for any specific disturbance
  - The rate of change of the position of the compressor operating point has been proven to be an excellent predictor of the strength of the disturbance
  - This derivative calculation can be used to adjust the magnitude of the open loop response
Recycle Trip® Based on Derivative of $S_s$

**Recycle Trip® Response Calculation:**

$$ C = C_1 T_d \frac{d(S_s)}{dt} $$

where:
- $C$ = Actual step to the valve
- $C_1$ = Constant - also defines maximum step
- $T_d$ = Scaling constant
- $d(S_s)/dt$ = Rate of change of the operating point

**Benefits:**
- **Maximum protection**
  - No surge
  - No compressor damage
- **Minimum process disturbance**
  - No process trips

**Graphs:**
- **Medium disturbance**
- **Large disturbance**

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All rights reserved. No part of this document may be reproduced or transmitted in any form or by any means, electronic, mechanical, photocopying, recording, or otherwise, without prior written permission of CCC.
• When the operating point moves quickly towards the SCL, the rate of change (dS/dT) is used to dynamically increase the surge control margin
• PI controller reacts earlier
• Smaller steady state surge control margins can be used without sacrificing reliability
The Safety On® Response

Compressor can surge due to:

- Transmitter calibration shift
- Sticky antisurge valve or actuator
- Partially blocked antisurge valve or recycle line
- Unusually large process upset

Benefits of Safety On® response:

- Continuous surging is avoided
- Operators are alarmed about surge

SOL - Safety On® Line
SLL - Surge Limit Line
RTL - Recycle Trip® Line
SCL - Surge Control Line

If Operating Point crosses the Safety On® Line the compressor is in surge.

The Safety On® response shifts the SCL and the RTL to the right.

Additional safety or surge margin is added.

PI control and Recycle Trip® will stabilize the machine on the new SCL.
Automated Fallback Strategies

- Statistically over 75% of control loop problems originate from field devices
- The antisurge system should have fallback strategies to handle transmitter failures problems
- System should continuously monitor input validity
- Input problems/failures should initiate a fallback mode of operation without a machine/process trip

Benefits:
- Nuisance Trip Avoidance
- Latent Failure Alarm
- Increased Machine & Process Availability
Compressor Performance Control
Performance Control

- Also called:
  - Throughput control
  - Capacity control
  - Process control

- Matches the compressor throughput to current process requirements

- Typical Main Process Variables:
  - Suction pressure
  - Discharge pressure
  - Net flow to the user
Performance Control by Blow-off or Recycle

Notes:
- Most inefficient control method
- Regularly found in plant air systems
- Rare in other systems
- Not recommended
- Curve 2 represents:
  - Lower speed on variable speed systems
  - IGVs closed on variable geometry compressors
  - Inlet valve throttled on fixed speed compressors
**Performance Control by Discharge Throttling**

- Compressor operates in point A
  - Shaft power: \( q_r \)
  - Pressure: \( P_d \)

- Required power is \( P_1 \)
  - Pressure is controlled by pressure drop over valve
    - \( PIC - SP \)
    - Pressure loss across valve

- Opening of valve would reduce resistance to \( R_{process} \)
  - Lower resistance would require less speed and power
  - Curve 2
  - Power loss is \( P_1 - P_2 \)
  - Lower speed on variable speed systems
  - IGVs closed on variable geometry compressors
  - Inlet valve throttled on fixed speed compressors

**Notes:**
- Extremely inefficient (consumes approximately the same power for every load)
- Not recommended
- Lower speed on variable speed systems
- Inlet valve throttled on fixed speed compressors
- Rarely used
  - Curve 2 represents:
  - \( R_{process} + R_{valve} \)
Performance Control by Suction Throttling

- Inlet valve manipulates suction pressure $P_d$
- Changing suction pressure generates a family of curves
- Pressure is controlled by inlet valve position $R_{process}$
- Compressor operates in point A for given $R_{process}$
- Required power is $P_1$

Notes:
- Common on electric motor machines
- Much more efficient than discharge throttling
- Power consumed changes proportional to the load
- Throttle losses are across suction valve

Shaft power vs. $q_r^2$

$P_1$

Suction valve open
Suction valve throttled

Notes:
- Common on electric motor machines
- Much more efficient than discharge throttling
- Power consumed changes proportional to the load
- Throttle losses are across suction valve
Performance Control by Adjustable Guide Vanes

- Change of guide vanes angle results in different compressor geometry.
- Different geometry means different performance curve.
- Pressure is controlled by inlet guide vane position.
- Compressor operates in point A for given R_process.
- Required power is P_1.

Notes:
- Improved turndown
- More efficient than suction throttling
- Power consumed is proportional to the load
- Power loss on inlet throttling is eliminated
Performance Control by Speed Variation

Notes:
- Most efficient: $(\text{Power} \gg f(N)^3)$
- Steam turbine, gas turbine or variable speed electric motor
- Typically capital investment higher than with other systems
- No throttle losses
Control Loop Interaction
Control Loop Interactions

• An optimized turbomachinery control system should incorporate all continuous control aspects of the machine
• Each continuous control loop interacts with other “local” and active control loops
• Interactions can negatively affect:
  - Machinery protection
  - Process stability & protection
  - Energy efficiency
• A common, less than optimum solution
  - Performance (process) control in DCS
  - Independent turbine speed governor
  - Independent antisurge control system
Antisurge & Process Controllers will Interact

• The antisurge controller and the process controller(s) manipulate the same variable:
  - The operating point of the compressor
• Multiple control loops on a single PV will interact when both loops are active
  - The control action of each controller affects the other
• The controllers have different and sometimes conflicting objectives
• Interaction with the A/S system begins on the surge control line where control stability is most important
Common Methods of Coping with Loop Interactions

- **De-tune the loops to minimize interaction**
  
  Result - poor pressure control, large surge control margins and poor surge protection

- **Put one loop on manual, so interaction is not possible.** Operators will usually put the antisurge control in manual
  
  Result - no surge protection and often partially open antisurge valve

- **Decouple the interactions**
  
  Result - loop stability, minimized setpoint deviation, more effective surge protection, and smaller surge control margins
Interacting Process & Antisurge Loops

• As Rc rises, the PIC works to reduce throughput through inlet throttling or asking for lower speed
• The disturbance and the PIC control response moves the operating point toward surge
• Interaction starts at B as the antisurge controller starts to operate
• At B, the a/s controller begins to open the a/s valve reducing Rc and increasing flow
• PIC continues to push the OP towards surge, along the resistance curve
Control Loop Decoupling

- Using loop decoupling, the a/s controller asks the process controller to simultaneously increase speed or open the inlet device.
- The combination of the decoupled process control output and the a/s valve opening provides a much more directly counteractive response to the direction of the disturbance.

Loop interaction:
- Causes instability
- Increases the risk of surge
- Increases setpoint deviation
- Increases safe margin requirement
Loop Decoupling for Multiple Antisurge Controllers

Disturbance comes from the discharge side
- $P_{d,2}$ increases
- $P_{s,2}$ remains constant
- $R_{c,2}$ increases
- Section 2 moves towards surge

Antisurge controller UIC-2 will open the recycle valve to protect Section 2 against surge
- $P_{d,2}$ decreases
- $P_{s,2}$ increases
- $R_{c,2}$ decreases
- Section 2 moves away from surge
Opening the recycle valve on Section 2 causes $P_{s,2}$ to increase / Result:
- $P_{d,1}$ increases
- $R_{c,1}$ increases
- Section 1 moves towards surge

UIC-1 will open its recycle valve to protect Section 1 against surge
- $P_{d,1}$ decreases
- $P_{s,1}$ increases
- $R_{c,1}$ decreases
- Section 1 moves away from surge
Opening of recycle valve on Section 1 caused $P_{d,1}$ to decrease / Result:

- $P_{s,2}$ decreases
- $R_{c,2}$ increases
- Section 2 moves towards surge

The system is oscillating
- Slowing down controller tuning could lead to:
  - Increased risk of surge
    - Compressor damage
    - Process trips
  - Larger surge margins
    - Energy waste
Interacting antisurge control loops need to communicate in real time to coordinate their control actions.

When UIC-2 opens the recycle valve:
- Section 2 will be protected against surge
- Section 1 will be driven towards surge

How much Section 1 is driven towards surge depends on how much the recycle valve on Section 2 is opened.

The Δ output of UIC-2 is sent to UIC-1 to inform UIC-1 about the disturbance.

The same applies when Section 1 antisurge valve is opened first.

The decoupling signal should be conditional, based upon a/s deviation.
Advanced Loop Decoupling

• Control functions should be integrated for coordinated mode switching required during start-up, shutdown, & trip scenarios
• Decoupling is required to avoid negative interactions between antisurge & pressure control loops
• Decoupling is required between antisurge control loops to prevent instability & excess recycle
Recycling for Capacity Control
Designers have the choice of using the antisurge control valves or dedicated recycle valves for capacity control.

- **Advantages of dedicated valves include:**
  - Process control selection criteria can be separated from the antisurge valve selection criteria
  - Low-noise trim & erosion resistant trim can be applied to the capacity control valves only
  - Can be considered a redundant valve arrangement
  - Opening of the a/s valves can be alarmed as “abnormal operation”

- **Advantages of using the A/S valve(s) include:**
  - Reduced space requirements
  - Reduced weight (offshore issue)
  - Reduced installation costs (valves & piping)
Use of Common Recycle Valve for Capacity Control

XIC -- "S" Controller for Capacity Control
UIC -- Antisurge Controller
PIC -- Suction Pressure Controller
Adding the “S” Control Line to the Compressor Map

XIC uses a low-select on the DEV calculation(s) of the associated A/S controllers
Hot & Cold Recycle Valves

- “S” Controllers used on cold recycle lines/valves for capacity control requirements
  - Low noise & anti-corrosion trim can be added to these valves
- Hot recycle used for short bursts of flow when UIC control line is crossed & for extra “trip Cv” as needed
  - Operating point will settle out on the cold recycle line during low-load conditions
Loadsharing for Compressor Networks
Control System Objectives for Parallel Compressors

- Maintain precise control of the primary process variable (pressure or flow)
- Provide integrated and bumpless limiting variable control including surge control
- Optimally divide the load between the compressors in the network, while:
  - Minimizing the possibility of machine trips due to surge, vibration, over-speed, etc.
  - Maximizing production throughput
  - Minimizing energy consumption
  - Minimizing disturbances of starting and stopping individual compressors
Operators select desired speed setpoint for the base-load machine.
**Base Loading for Parallel Compressor Operation**

Compressors operate at the same $R_c$ since suction and discharge are common.

- Base compressor is fully loaded and runs without recycle while the other(s) absorb the load swings.

$$QP,1 + QP,2 = QP,1 + QP,2$$

**Notes:**
- Base loading can lead to inefficient operation.
- Base loading increases the risk of surging the swing compressor as it will take the worst of any disturbance.
- Base loading requires frequent operator intervention.
- Base loading is NOT recommended.

Where:
- $Q_p$ = Flow to process
- $Q_c$ = Total compressor flow
- $Q_c - Q_p$ = Recycle flow
“Equal Flow” Control for Parallel Compressor Operation

- Process control action is independent of the antisurge control
- Higher capital cost due to extra flow measurement devices (FMD)
- Higher energy costs due to permanent pressure loss across FMDs
“Equal Flow” Control for Parallel Compressor Operation

Compressor 1

\[ Q_{P,1} = Q_{P,2} \]

Compressor 2

Notes:
- Requires additional capital investment in FMDs
- Requires additional energy due to permanent pressure loss across FMDs
- Poor pressure control due to positive feedback in control system (see next)
- Equal flow division is NOT recommended

where:
- \( Q_P \) = Flow to process
- \( Q_C \) = Total compressor flow
- \( Q_C - Q_P \) = Recycle flow
In a classic cascade control scheme, the slave controller needs to be tuned 4-5 times faster than the master controller.

In steady state operation, the machine is operating at the intersection of:
- Resistance Curve
- Speed Curve
- PIC-1 Setpoint
- FIC-1 Setpoint

The Master Pressure Controller (PIC-1) provides the remote setpoint for Flow Controller (FIC-1), which sets the speed setpoint for SIC-1.

In steady state operation, the machine is operating at the intersection of the:
- Resistance Curve
- Speed Curve
- PIC-1 Setpoint
- FIC-1 Setpoint

“Equal Flow” Control Dynamic Control Problem
Scenario: A process disturbance causes the resistance curve to move from $R_1$ to $R_2$, shifting operating point to $B$.

- Since the PIC-1 loop gain is slow, its output (FIC-1's setpoint) moves slowly.
- FIC-1 reacts quickly trying to maintain its setpoint, speeding up the machine to point $C$ at speed $N_2$.
- PIC-1 slowly reduces FIC-1's setpoint and the primary process control variable (pressure) setpoint is finally restored at speed $N_3$.

Notes:
- Can cause significant instability near surge.
- Results in unnecessary process setpoint deviation.
- Traditional pressure to flow cascade systems are NOT recommended.
Equidistant Loadsharing

- LSICs get their PV from their companion A/S controller and their setpoint from the MPIC
- All controllers should coordinate their control responses in real-time
Equidistant Loadsharing

- Machines operate at same \( R_c \)
- The DEV is a dimensionless number representing the relative distance between the operating point and the SCL
- Machines are kept at the same DEV which may correspond to different speeds and flow rates
- Recycle will only start when all machines are on their SCL
- Since DEV is dimensionless, all sorts and sizes of machines can be operated together: large, small, ST & motor driven

\[ \text{DEV}_1 = \text{DEV}_2 \]
\[ Q_1 \neq Q_2 \]
\[ N_1 \neq N_2 \]

Notes:
- Maximum system turndown without recycle or blow-off
- Minimizes the risk of surge since all machines’ (recycle valves) absorb part of the disturbance
- Automatically adapts to different types of compressors
- CCC patented algorithm
The Primary Response

- Master controller (MPIC) controls the main Process Variable (PV) via its PID control block
- The output of the MPIC PID goes to the primary response block in the loadsharing controller
- In the primary response block, the controller checks if the machine is close to the SCL:
  - Yes: don’t reduce compressor throughput
  - No: reduce compressor throughput as necessary
- Apply loadsharing gain $M_0$
- The primary loop gain of MPIC goes directly to the performance control element via the primary response block
The Primary Response

- In order to check if the machine is close to the SCL, the LSIC primary response block needs the A/S DEV
- The DEV is reported by the UIC
- When the machine is close to the SCL, MPIC output changes will be redirected from the LSIC to the UIC
- When DEV < or = 0, further reduction of machine throughput is handled through modulation of the antisurge valve

Diagram:
- MPIC
  - PID
  - PV
- UIC
- LSIC
- Primary Response
  - DEV
  - To antisurge valve
In order to maintain equal distance from surge, each compressor needs to be kept at the same antisurge DEV. The antisurge controller reports its current DEV to the load balancing block in the loadsharing controller. This reported DEV becomes the Process Variable (PV) for the load balancing PID loop in the LSIC. The loadsharing controller also reports this DEV to the MPIC.
• Other loadsharing controllers also report their DEV to the master controller to be averaged
• The MPIC sends the average DEV to all LSICs as the set point for all load balancing blocks
• The load balancing block works to carefully balance the DEVs on all running compressors
• Its output is added to the total output to the performance control element