



46TH TURBOMACHINERY & 33RD PUMP SYMPOSIA
HOUSTON, TEXAS | DECEMBER 11-14, 2017
GEORGE R. BROWN CONVENTION CENTER

IMPROVING LNG FACILITY RELIABILITY AND OPERABILITY VIA OEM INTEGRATED COMPRESSOR CONTROLS

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ABSTRACT

This paper will discuss a multi-string LNG compressor system's availability and reliability impact using an integrated compressor control and protection system. Leveraging knowledge gathered with operating LNG plants, several operability, system availability and reliability improving opportunities were identified and implemented in a new grass roots facility. The paper will present the execution of the control systems project from the early concept select phase to, front-end to detail engineering, testing, commission and start-up. It will also show the close integration for the control system design with the plant EPC activities.



INTRODUCTION

During the design phase of the LNG project, the Plant Operator described previous experiences where compressors would trip due to surge events in process upsets. These compressor trips would prolong facility upset conditions, interrupt production and expose equipment to unnecessary increased transient events. Also, on the restart of a compressor the process was negatively impacted due to excessive recycle while placing it into service and load sharing between compressors in similar service was difficult often resulting in compressor trips. The Plant Operator also required a control system which manages all facets of Turbine / Compressor control and protection from a single software / hardware platform. The single platform would also allow Operators to View and Control each Turbine / Compression string from any single workstation in the compressor control room as well as the central control room. Besides operating more efficiently due to the common hardware / software platform integrated with compressor and turbine performance, the hardware interchangeability also raises opportunities advantageous to long term maintenance strategies of the control systems.

This challenge, including providing an interface for the Operations staff integrated with the process design, was specified to the turbine and compressor OEM as an engineering and design requirement. The OEM responded by providing an integrated control system, preventing the Operations staff from managing complex manual procedures. Stipulated operating cases included normal operation and off-design operating modes, such as start and stop transients, large feed-rate changes up to and including plant full to half feed rate and reverse transitions.

To address compressor availability, the OEM provided a strategy to manage instrumentation failure via an on-line calculation of failed instruments using compressor analytics, using its aerodynamic codes and “as tested” compressor performance maps. Compressor performance calculations were also provided for compressors with process side streams to increase accuracy of the operating point calculations. This method allows a more efficient usage of the compressor optimizing the control of anti-surge valves.

During the design phase of the project, operability scenarios were selected and simulated by “software in the loop” and “hardware in the loop” methods. Moreover, the compressor control system was integrated into the Operator Training System (OTS) where the control system was tested and tuned with the process.

At commissioning phase, a dedicated OEM controls commissioning procedure was executed to tune each regulator by identifying system open loop responses. Once configured, each regulator was then validated in closed loop. Commissioning tests were designed to be operated during LNG production phase with negligible to no impact on plant production.

It should also be noted, the improved stability of the process allowed the compressor operating point to surge point design margins to be reduced. The resulting reduced recycle flow, provided for a measurable higher production efficiency, while being in a safe and controlled operating environment.

LNG PLANT BASICS AND CONTROL OBJECTIVES

An LNG refrigeration plant is a complex system that transforms feed gas at near ambient temperature to liquid at about -160°C (-256°F). The liquid is stored with a volume reduction gain of about 600 and transferred to LNG carriers to be transported to destination. The feed gas temperature decreases through heat exchangers and cold boxes by transferring heat to plant refrigerants. Refrigerants are continuously boiling in dedicated equipment and gas produced flows through closed loop circuits pushed by compression strings. Figure 1 shows a simplified flow diagram of the LNG technology discussed in this paper. However, the same compressor control strategy and design may be applied to refrigeration compressor utilized in other technologies, or applications for that matter. Variable feed rate is managed by adapting refrigerant flow rate through variable speed compressor drivers.

Maintaining stability of operation is one of the important objectives of the plant operation team, which mitigates instabilities that could adversely impact production rates. Stability is assured when boiling refrigerants are kept at the constant temperature designed so that feed gas can liquefy. From saturation curve (see Figure 2), the temperature requirement is translated to pressure requirement. From this observation originates the main objective of process automatic control of compression strings that is to maintain a constant string suction header pressure. Quickly responding to process induced disturbances, costly negative production impacts may be avoided.

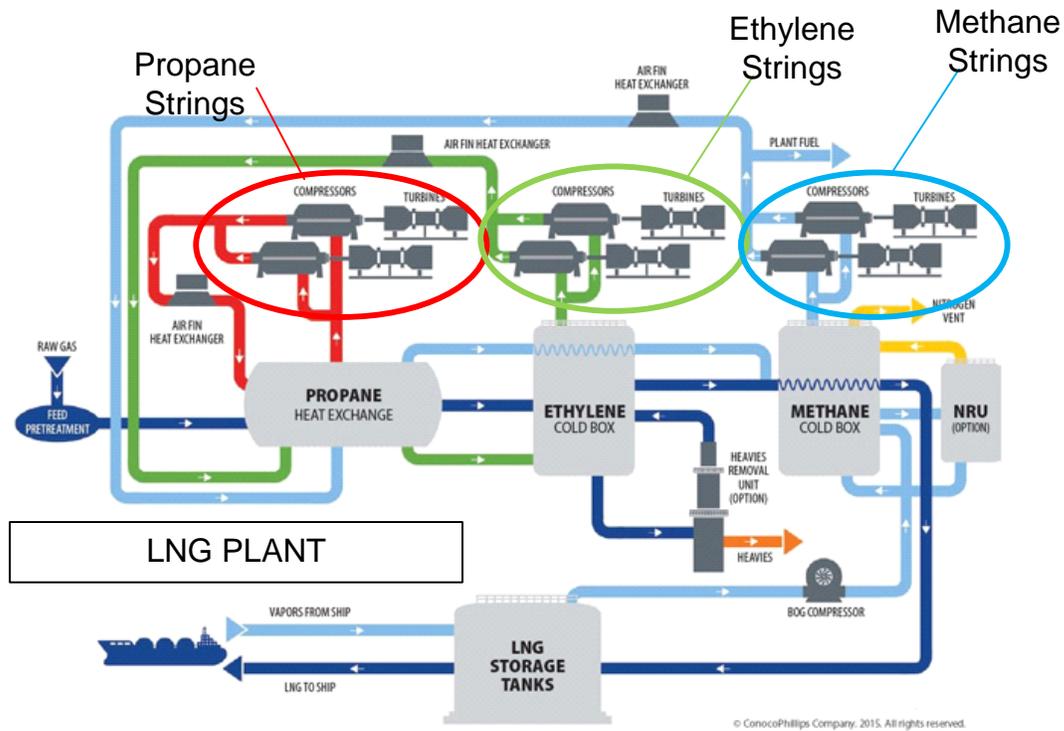


Figure 1 – LNG Plant simplified flow diagram

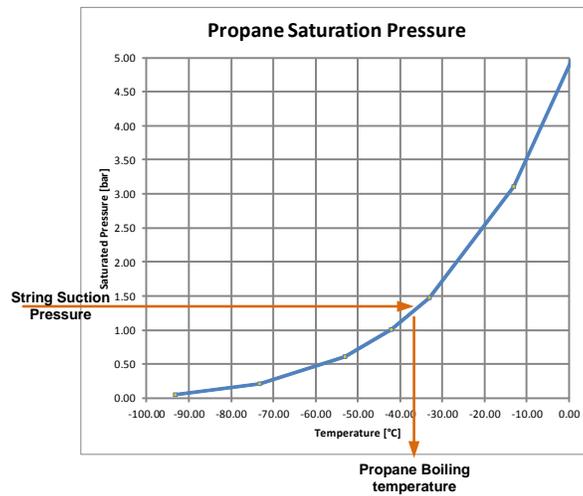


Figure 2 – Saturation curve pressure/temperature relationship

Schematizing compressor strings common suction header as a volume which flow input corresponds to boiling gas flow and flow output is managed by variable compressor speed (see Figure 3), constant pressure within suction header requires balancing mass flows so that:

$$G_{in} [kg/s] = G1_{out} [kg/s] + G2_{out} [kg/s]$$

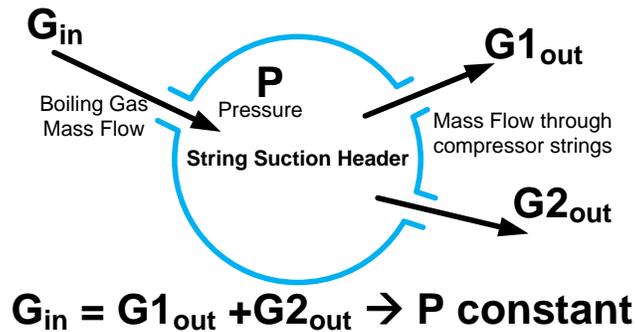


Figure 3 – String Mass Balance and string suction pressure

If the equality above is not met, a pressure drift proportional to mass balancing error occurs, causing loss of refrigerant temperature with effects or consequences propagating to other plant or process areas.

While refrigerant boiling mass flow G_{in} is influenced by process variables that are linked to heat transferred to the refrigerant itself, the flows through compression strings ($G1_{in}$ and $G2_{in}$) are controlled mainly by adjusting the compressor speed. Balancing mass flows requires a continuous adjustment of compressor speed to maintain a set suction pressure.

Introducing the concept of compressor operating envelope, the above perspective is somewhat complicated by the fact that the compressor speed cannot be decreased indefinitely due to the compressor surge phenomenon that occurs if operating compressor on the left of surge line, below its required minimum flow, showed in Figure 4. To permit operating the compressor below minimum required flow, hence protecting the compressor from destructive surge events, a flow recycle line with a control valve is installed in parallel with compressor. The valve is referred to as anti-surge valve (ASV). The string flow can be decreased by speed when operating the compressor within its operating envelope and by combined action of speed and ASV when operating at the operating envelope left boundary. If driver minimum operable speed is reached, string flow can be further decreased by opening of ASV, until zero flow condition. On the other side, string flow can be increased until max speed or compressor choke line is encountered. Usually, driver power limit intervenes before maximum speed is reached.

Anti-surge valve opening reduces efficiency of operation because part of the power is consumed when recycling the gas. Because of its negative impact on system efficiency, recycling operation is to be optimized, limiting the ASV opening to manage low compressor flow scenarios or also during fast process disturbances due to achievable high rate of response in controlling suction pressure through ASV.

Looking again to the mass balance equation, we note that there are multiple solutions in the choice of string flows $G1$ and $G2$ so that their sum verifies the balance equality of Figure 3: this degree of freedom is used to maximize the efficiency optimizing the opening of any anti-surge valve. The efficiency objective is achieved by controlling parallel compressors to work at same “margin to recycling” to reduce the opportunity of opening the anti-surge valves that is the main cause of system inefficiency. As the recycling is needed at left compressor boundary limits, operating point margin to recycling can be conveniently defined through a combination of speed and flow measurement, resulting in the contour plot indicated in Figure 5.

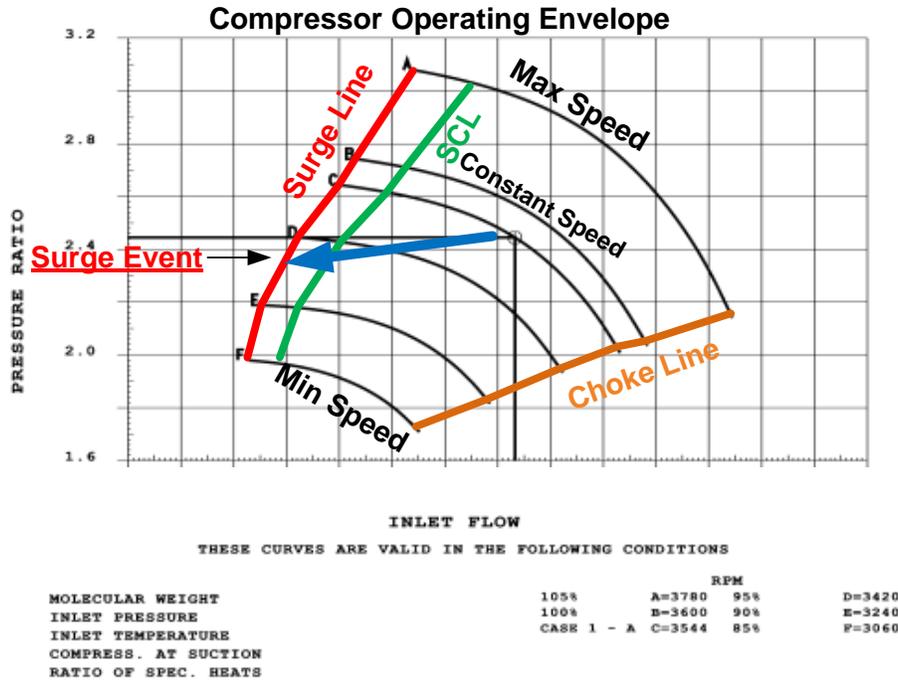


Figure 4 – Typical Compressor Operating Envelope

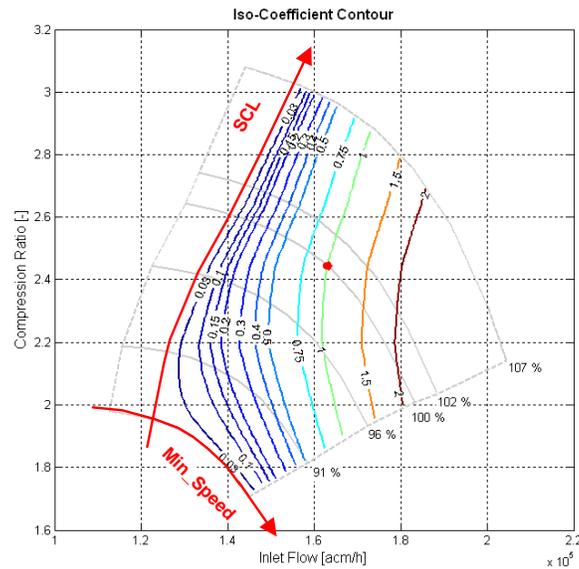


Figure 5 – Distance to recycling definition



Other process variables are not required to be kept at a fixed value but is important to maintain them within an acceptable range. As an example, string discharge pressure needs to be maintained below a limit value, above which a string process trip value is placed. When approaching that limit, string controlling variables cannot be operated just to regulate suction pressure as described above but all control effort switches to the limiting control to avoid a string “trip”, meaning an unplanned stop of string driver.

From the above notes, compressor strings main operation requirements are summarized as follows:

- 1- Regulate compressor string suction pressures at a given set-point.
- 2- Regulate compressors within their operating envelope
- 3- Load shared through compressor strings to maximize efficiency of operation
- 4- Limiting pressure control on compressor string discharge
- 5- Minimizing compressors surge margin

Of course, such requirements are not practical to be managed manually by plant operators because multiple actions need to be done simultaneously especially during plant transients.

Requirements in the above list are objectives and constraints for automatic control system, aiming to assist and optimize plant operations.

The compressor string diagram of Figure 6 shows control inputs to calculate the compressor operating point, to calculate the deviation of measured string suction pressure to a desired set-point, to calculate margins to limiting actions at discharge. Also, controlling variables, speed and ASV, are shown.

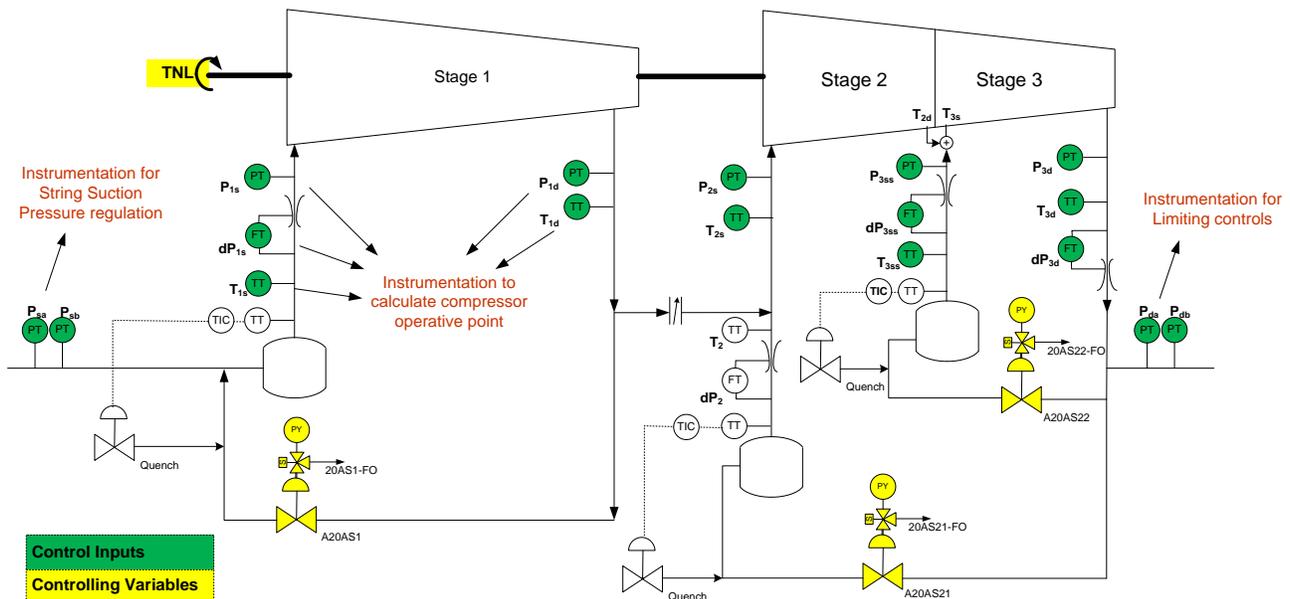


Figure 6 – Typical Refrigeration Compressors String Diagram



BASELINE REPORT

Based on lessons learned from operating facilities from past experiences, plant operation personnel identified system design improvement opportunities. They were collected in a design specification or guideline and communicated to the CC OEM early in the project design phase.

Failures reported spans following items:

- Trips induced by process (“process trips”) due to in-correct management of plant upsets (i.e. feed loss)
- Inadequate process transient management, including loss of a single and or multiple compressor string
- Complex manual operation to load/unload strings
- Lack of adequate corrective actions in case of loss of signal (trip and valve trip reductions)
- Poor flow control through ASV due to large C_V selection

Close and early collaboration between engineering and operation with the compressor OEM was the key element to address operability issues, by firstly analyzing root causes of unwanted control system responses and operations, then defining detailed and measurable control requirements which had to be validated by a rigorous testing campaign of the controllers and integrated with the OTS (Operating Training Simulators).

AUTOMATIC CONTROLS DESIGN

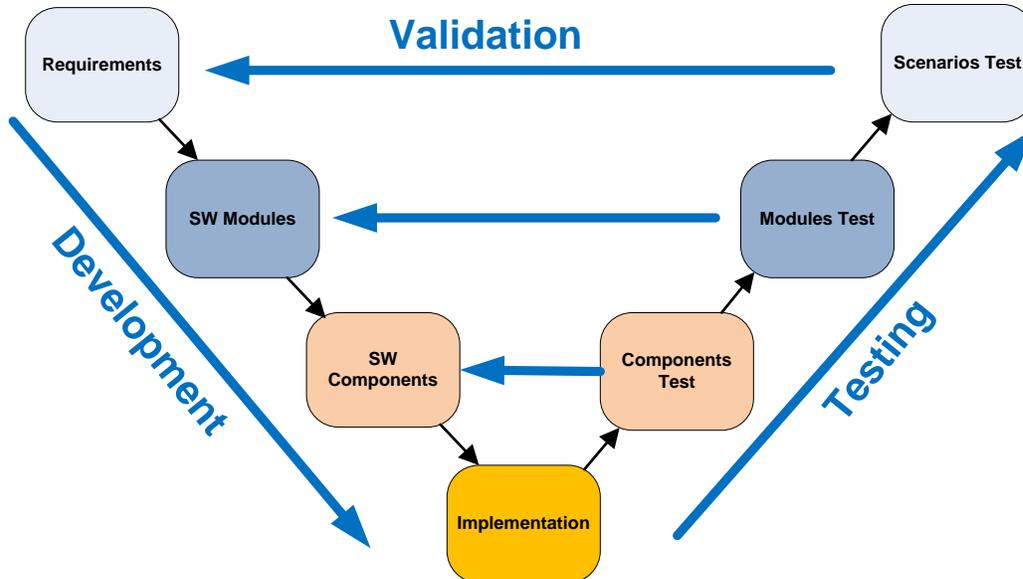


Figure 7 – Controls design process

Starting from the requirement analysis, any design is an iterative process as depicted in Figure 7. The process is comprising of a concept select, development, engineering, testing and validation phases. At development, requirements are translated in software modules, each one split in components, then implemented. The test and validation phases requires to develop a simulation environment in which the plants behavior can be digitally replicated using known physical equivalences. Strictly following the design process, defined opportunities are addressed and enhanced control solutions implemented and tested. It is critical that results are being evaluated against the set of design and operational objectives. If required, the process is repeated till a solution meeting all operational objectives are met and verified.



In fact, the conceptually simple process pressure regulation hides complexity that requires to be properly addressed by the control system. Early identified opportunities possibly delivering an enhanced control system included operational matters related to:

- a) Regulators interaction
- b) ASV position to compressor flow dead time
- c) ASV actuator stiction,
- d) Side stream compressor operating point calculation
- e) Process control availability in case of compressor instrumentation fault

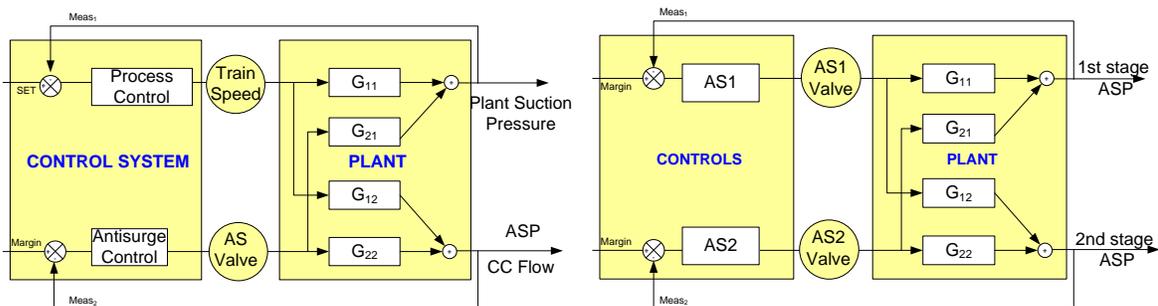


Figure 8 – Typical process SISO regulators couplings scheme

a) Regulators Interaction:

Multiple controlled variables within the process are typically regulated by Single Input Single Output (SISO) controllers: each controller regulates a single controlled variable actuating a single controlling variable. Interaction between SISO regulators occurs when a single controlling variable impacts multiple controlled variables. In this case, the system to be controlled is described as “coupled”. For example, process control actuates speed to regulate suction pressure, but it also affects compressor flow mainly regulated by anti-surge control (see Figure 8). When using SISO regulators to manage a coupled system, coupling effects are managed as control disturbances: this worsen dynamic response towards long transients until instability.

Many coupled conditions have been identified at the simulation stage and then observed in the real plant. As an example, Figure 9 shows the result of opening the second stage ASV of string 1 with a 5% step when both strings are running. It is noted that due to the connections of the two strings, the compressor flows change not only in the second stage of the string with the stepping valve, but also at the second parallel string.

In configuration of Figure 10, the coupling is so high to be prevalent: the same ASV step is given when the parallel string is at higher speed. The ensuing compressor flow change is more evident in the parallel string. In this scenario, the controlled variable is marginally affected by controlling variable, reducing system controllability. The same effect has been also observed at same string speed in a startup scenario, characterized by a cold and a hot parallel string due to different opening of ASV. The explication is straightforward introducing the concept of “corrected speed” (N_c), for which compressors maintain similarity when running at same corrected speed. Corrected speed is calculated as follows in case of compressor processing a generic gas:

$$N_c = N \frac{\sqrt{k_v \frac{Z_s RT_s}{M_w}}_{@ design}}{\sqrt{k_v \frac{Z_s RT_s}{M_w}}}$$

At startup, due to the difference on suction temperature, the two parallel compressors will run at different corrected speed, so behaving differently also if running at same physical speed.

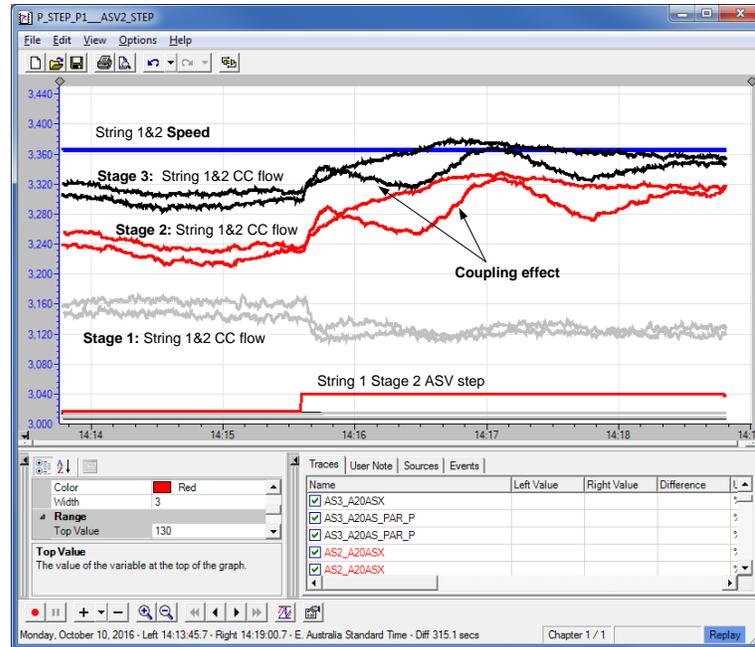


Figure 9 –Coupling open loop test: ASV first stage position interfere with all compressors of both strings

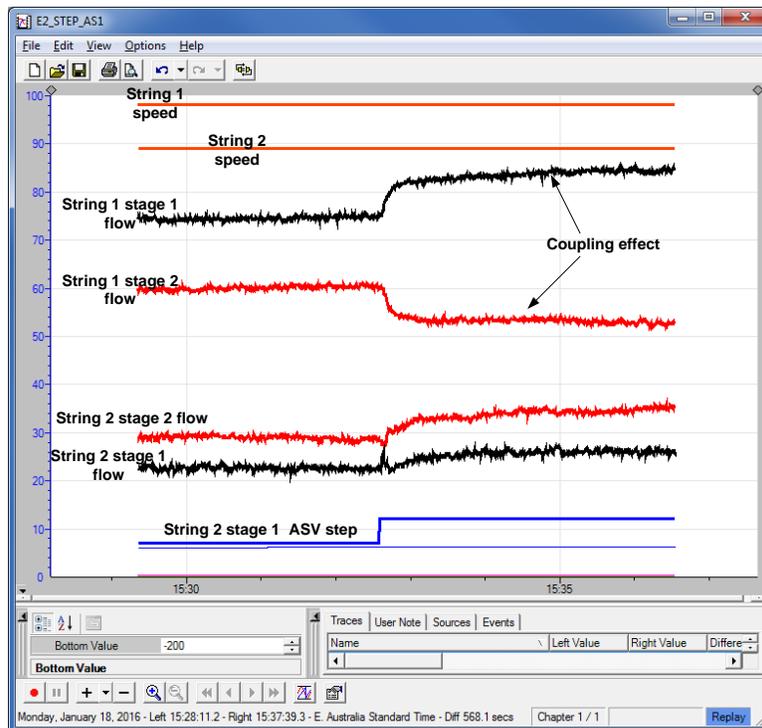


Figure 10 – In case of unbalanced speed, ASV has more effect on parallel string (trended at commissioning)



As a clear result, regulator interaction needs to be managed by control system. During the design stage, following control strategies were adopted:

- Feed-forward de-coupling
- Matching parallel compressors anti-surge openings
- Load control SIMO (Single Input Multiple Output) strategy

Decoupling through feedforward actions has been implemented as in Figure 11 where D_{12} and D_{21} are determined by process identification test to cancel the coupling effect:

$$D_{12} * G_{22} + G_{12} = 0 \quad \rightarrow \quad D_{12} = -\frac{G_{12}}{G_{22}}$$

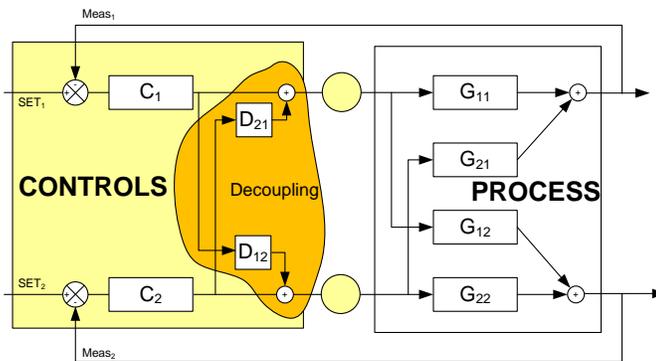


Figure 11 –Decoupling through feedforward

In case of different corrected string speeds, observed couplings reduces controllability of SISO loops (see Figure 10). To avoid this scenario, parallel ASV are kept aligned by a dedicated Parallel Override logic (POV) shown in Figure 18.

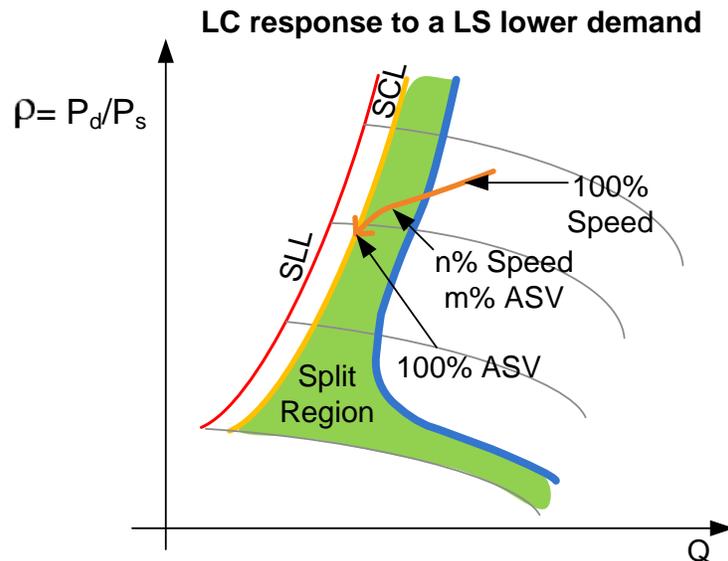


Figure 12 – Speed/ASV split concept



As noted previously, the string's suction pressure is affected both by compressor string speed and anti-surge valves (see Figure 8). From the efficiency aspect, it is preferable control by speed but, when approaching the compressor minimum flow limit, operating speed cannot be decreased further and an action by the ASV is needed. Controlling variable transition from speed to ASV is given gradually depending on margin to recycling value as shown in Figure 12: flow demand decreasing is initially managed by speed only then, in split region, the demand is gradually given to ASV and, on SCL, the demand is completely given to ASV. Regulation becomes SIMO, with each part being tuned separately. This regulation switching strategy helps to avoid suction pressure and antisurge regulators interaction with positive effect on control stability.

b) ASV position to CC Flow dead time

Responding with the ASV to process upsets requires fast action by ASV. Typically, a dead time between ASV positioning and corresponding compressor flow is observed. This is mainly due to actuator characteristic and flow transport delay. A dead time value of 1s (see Figure 13) impacts stability when pushing regulator gain to speed up the response. To manage the dead time, the regulator design has been modified as shown in Figure 14, introducing a Smith Predictor scheme modified by a process model to calculate the compressor corrected flow. The model output is used to calculate the difference between delayed (through a Padé filter) and non-delayed model response. The difference then compensates the calculated compressor corrected flow from field measurements that includes the process delay. This scheme allows to work with higher regulator gains without affecting stability.

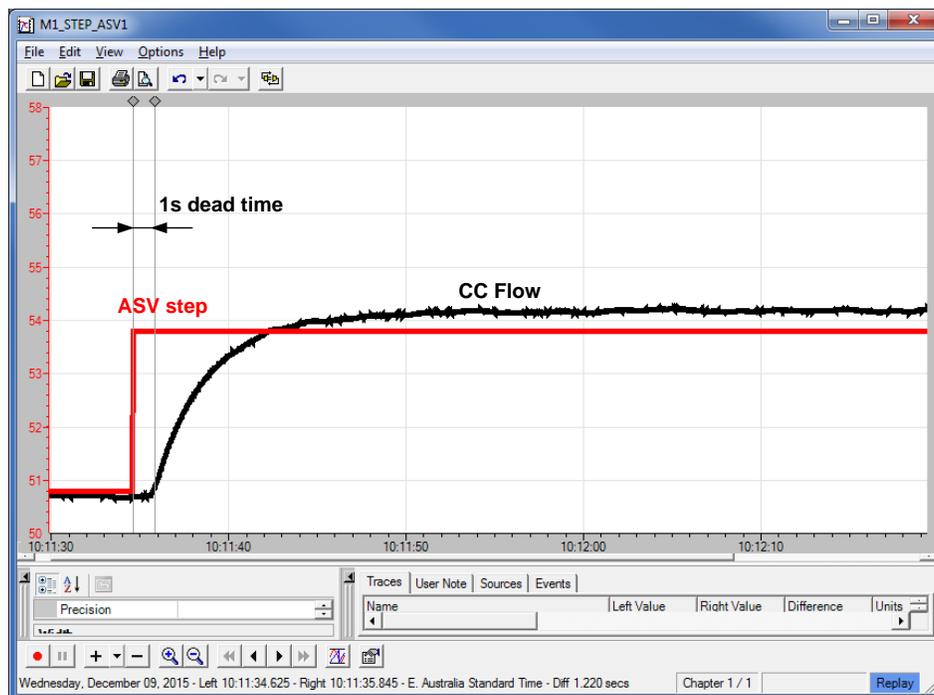


Figure 13 – ASV to compressor flow dead time trended at commissioning

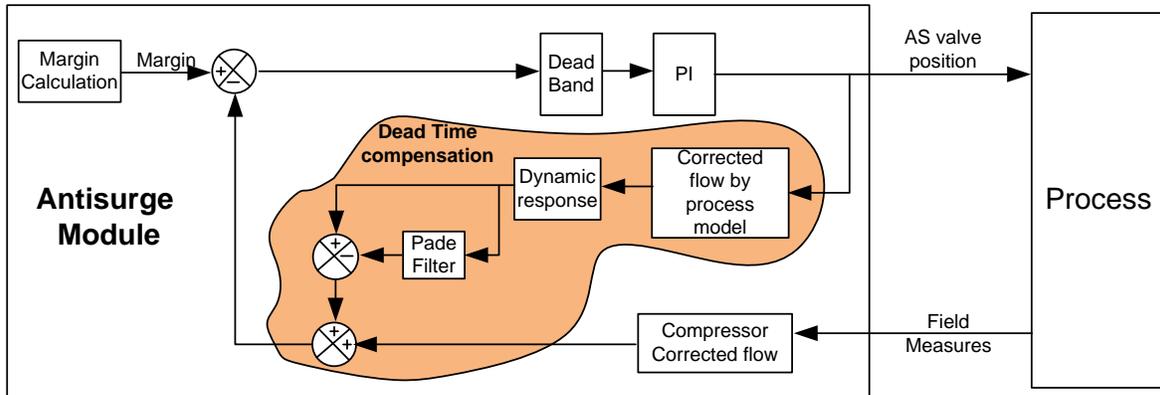


Figure 14 –Dead time management through Smith Predictor

c) ASV actuator stiction.

ASV actuator stiction, detected by jerky movements as shown in Figure 15, reduces controllability: the ASV movement is not smooth and controlled compressor flow through ASV never reaches a steady state condition. This effect is eliminated or greatly reduced by properly selecting valves with minimum value of CV then introducing a regulator dead band that is properly tuned.

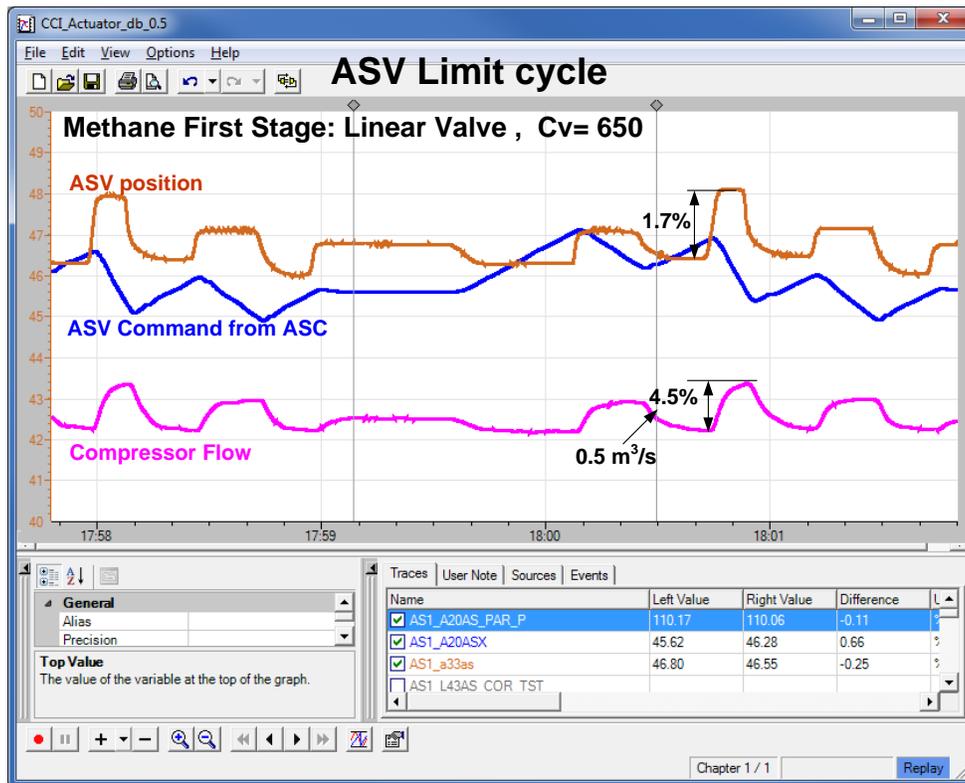


Figure 15 – Effect on compressor flow regulation of ASV stiction



d) Side stream compressors operation

To identify a compressor operating point position within operating envelope, corrected flow m_c is calculated through field measurement. In case of compressor flow measurement at suction, corrected flow is calculated as follows:

$$m_c|_{FE@ suction} = \frac{Q_{vs}}{a_s} = \frac{K_{FE}}{\sqrt{k_{vs}}} \sqrt{\frac{dp_s}{P_s}}$$

In case of a compressor with one or more side streams as shown in Figure 16, the calculation of corrected flow becomes more complex being also dependent by non-measured quantities. As an example, second stage corrected flow of Figure 16 is the following:

$$m_c|_{Stage2} = \frac{Q_{v2s}}{a_{2s}} = \frac{K_{FE}}{\sqrt{k_{v2s}}} \sqrt{\frac{dp_{2d}}{P_{2s}} \frac{P_{2d}}{P_{2s}} \frac{T_{2s}}{T_{2d}} \frac{Z_{2s}(P_{2s}, T_{2s})}{Z_{2d}(P_{2d}, T_{2d})}}$$

Noting that second stage suction temperature (T_{2s}) is not measurable, previous methods to calculate the operating point used rough approximations for suction temperature and for compressibility. The calculation was done on conservative basis with respect to compressor surge protection, selecting suction temperature to minimize calculated corrected flow. Being a conservative calculation, this caused unneeded opening of ASV so resulting in a loss of string efficiency.

The method has been updated to increase accuracy of corrected mass flow calculation by calculating suction temperature through compressor analytics derived by precise knowledge of compressor behavior and calculating compressibility ratio through known table values.

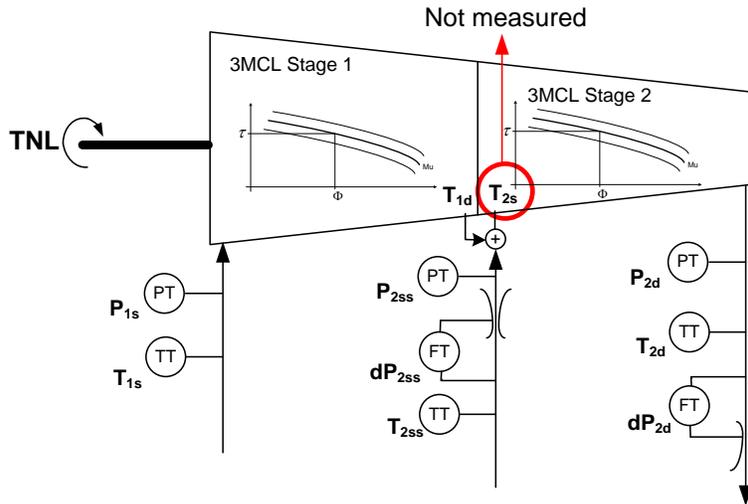


Figure 16 –Compressor with side streams: T_{2s} is calculated through compressor analytics

e) Process control availability at compressor instrumentation fault

In case of an instrumentation failure, the compressor operating point within the operating envelope cannot be identified. Non-dimensional compressor analytics are used to replace a faulty compressor measurement. The virtual variable approach supports increasing system availability, particularly in case of a single instrument configuration (single point of failure). This method improves the system accuracy and reliability when compared to the more common practice of “freezing” all control output at their current values until a steady condition is detected through the healthy measurements. In fact, in case of not perfectly steady state conditions, the default is to open the ASV to protect the compressor, hence sacrificing process optimization. As an example, Figure 17 reports the case of compressor discharge pressure instrument failure. The worst-case method maximizes the compressor pressure ratio using a maximum discharge pressure with the consequence to overcome surge line that cause a full opening of ASV with severe process consequences. The worst-case scenario is avoided using the “virtual” instrument provided by compressor analytics.

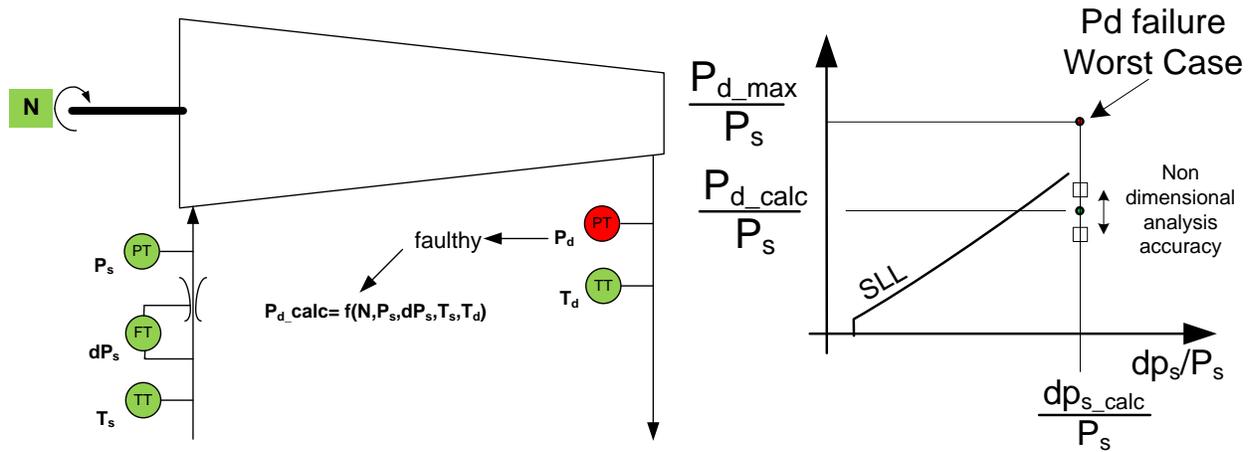


Figure 17 –Example of fallback in case of an CC instrument failure

CONTROLS ARCHITECTURE

At conclusion of the control system design phase a modular architecture as shown in Figure 18 has been selected to fulfill shared requirements.

Following four modules were identified (see Figure 18 for interconnections):

Process Control Module:

This service level module manages the service set-points, translating the process variables into compressor string operating point requests.

Load Sharing Module:

Receives the demand from the process controller and translate it to compressor string control, also considering startup, shutdown, balancing within the compressor strings, loading and unloading transient requirements (see Figure 19).

Load Control Module:

Translates compressor string process demands to ASV and speed requests (Figure 20).

Anti-surge Module:

Sketched in Figure 14, protects the compressor from surge through the ASV's. If within compressor operating envelope, the module selects the demand for the compressor string control.

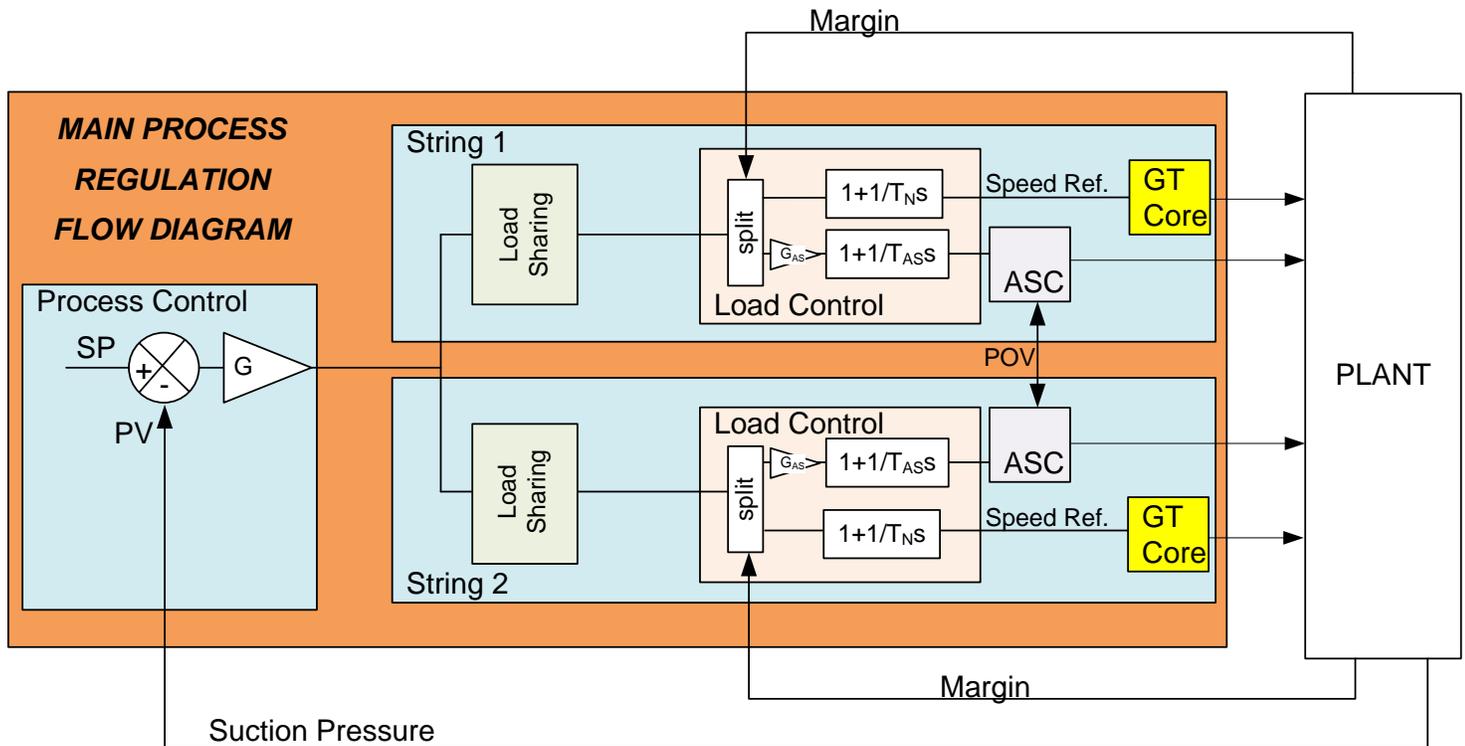


Figure 18 – Controls modular architecture

Regulators have been selected based on identified system responses. As step response to speed and ASV are both well described as a first order system, a PI regulator was selected. In fact, a PI regulator is preferred solution in case of a first order system to design the closed loop response as a first order transfer function which its time constant (α) is the design parameter. The tuning of PI parameters (gain and integral time) are calculated as function of open loop first order system parameters (gain and time constant) and the design parameter α . Also, the design parameter α is used to calculate the disturbance rejection capability of the closed loop system.

The regulator selection was also influenced by the simplicity of tuning through a single design parameter α and system identification tests. This helped to design the commissioning through a procedure that included tests and validations.

In the control scheme of Figure 18, the process control PI transfer function is implemented distributing it on multiple modules: the proportional part is implemented in process control module and the integral part is implemented in load control module. Note that calculating the overall transfer function of the two series terms, the classical form of PI (Proportional Integral) controller transfer function is obtained:

$$PI = G \left(1 + \frac{1}{T_i \cdot s} \right)$$

The load sharing module (see Figure 19) equally divides the process control incremental demand ($\Delta\text{FLOW_IN}$) through active compressor to maintain same control response independently by the number of active strings. Then sums various biases used in different phases of operation:

- during loading phase, identified by the string check valve closed, an additional bias is defined to accelerate the compressor to “exporting” (or loaded) state.
- Equalization bias is given to align parallel compressor strings. Strings alignment use the property that the sum of equalization biases through parallel strings is zero to not interfere with the main regulation.



- Unloading bias. Used by the operator to unload and segregate a string from the system without affecting the main process control. It works as opposite of equalization but still using the “zero sum” property.
- Trip Recovery: adds a bias to recover process regulation in case of the parallel string trip.
- Not Active Recovery: used when parallel compressor cannot manage main regulator bias due to max power limit intervention.

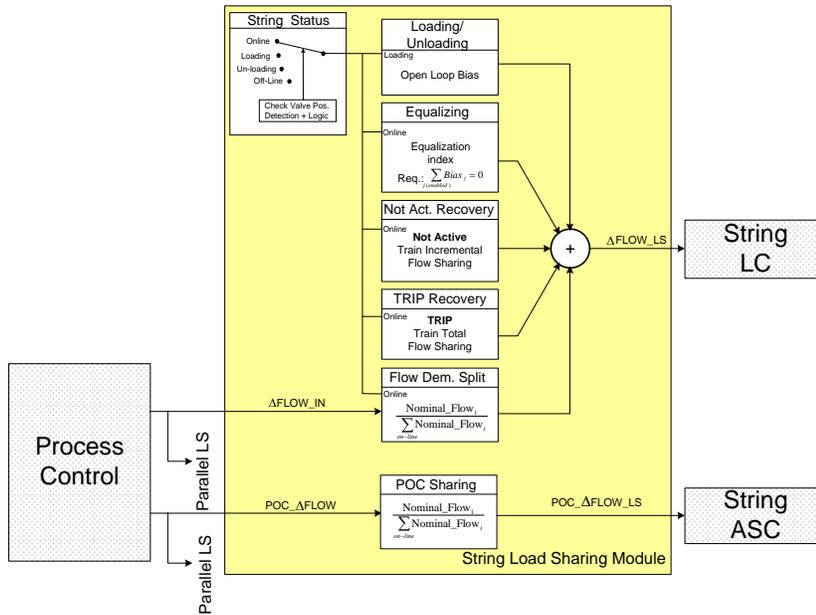


Figure 19 – Load Sharing Module

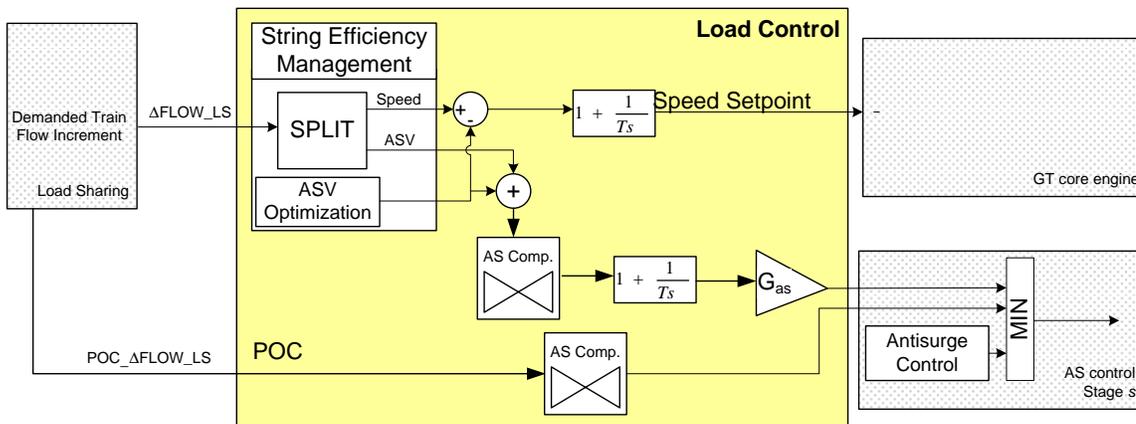


Figure 20 – Load Control module

The Load Control Module (Figure 20) manages the output of process regulation ($\Delta FLOW_LS$), string speed and ASV position, through the “split” logic to minimize coupling issues as described previously (see Figure 12). Another part implemented in load control module is the ASV optimization function that in case of a steady state condition with ASV not closed, it decreases the speed and closes ASV until ASV is fully closed (or on SCL). Also, command given to ASV is compensated through ASV stroke-flow characteristic.



Process control, load sharing and load control manage also an additional control loop (POC) that intervenes to limit controlled variables overshoot acting on ASV only.

Use of Simulations

The simulations used during the design stages start with PC based simulations (see Figure 21) as functional testing to verify defined engineering and operational specifications are implemented and are functional. The dry rig based simulations linking the job control hardware to the plant simulator or OTS (Operator Training Simulator) is the final stage of the simulated testing program (see Figure 22). Testing with the OTS is also used to train the plant operators.

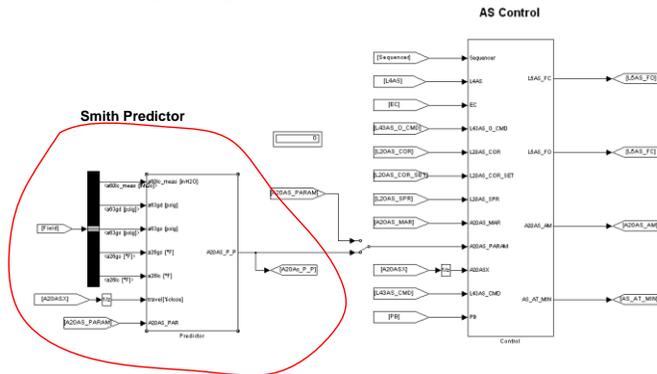


Figure 21 –Simulation in PC environment example

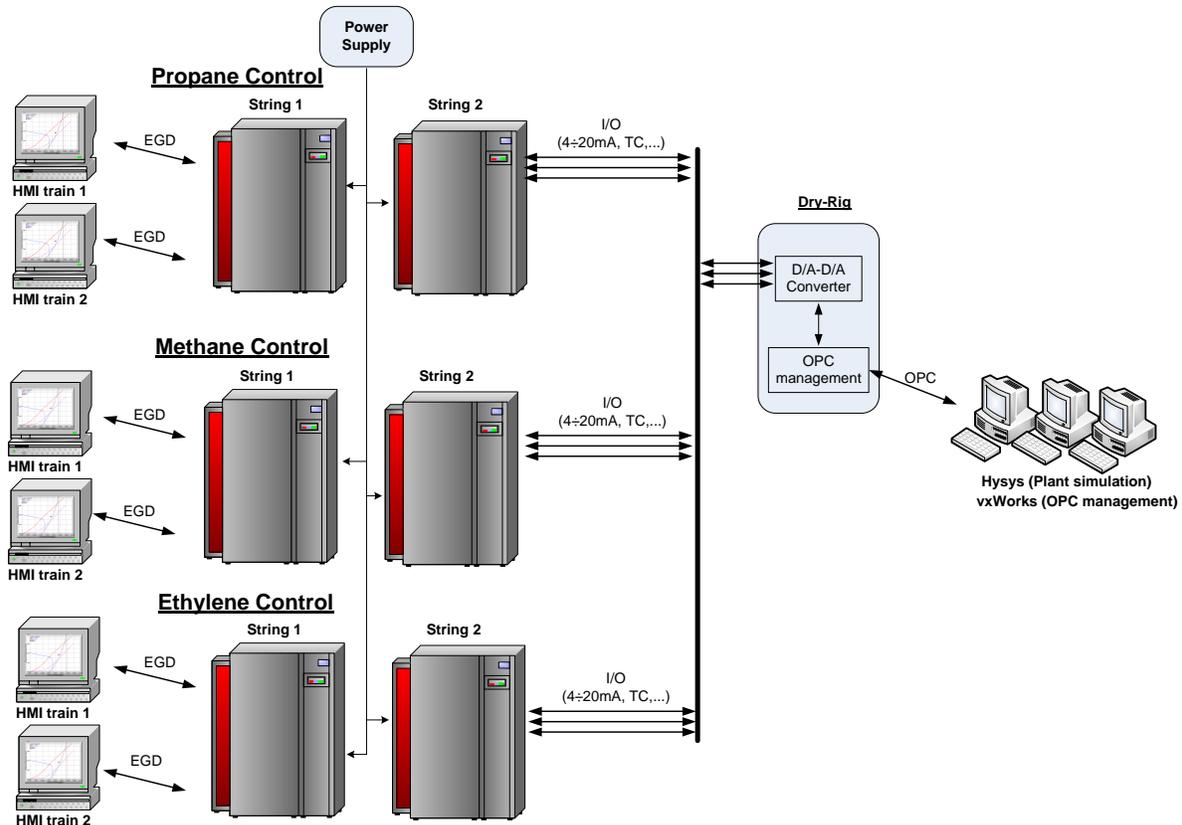


Figure 22 – Hardware in the loop test configuration



CONTROLS COMMISSIONING

Process control commissioning starts after completion of the usual gas turbine driver and compressor commissioning, when plant is operational and feed gas is introduced. With production being imminent, this part is critical because commissioning test could slow-down production rate. To avoid costly delays, commissioning tests need to be designed early. They should be transparent to production and operators and executed with operation personnel supervision.

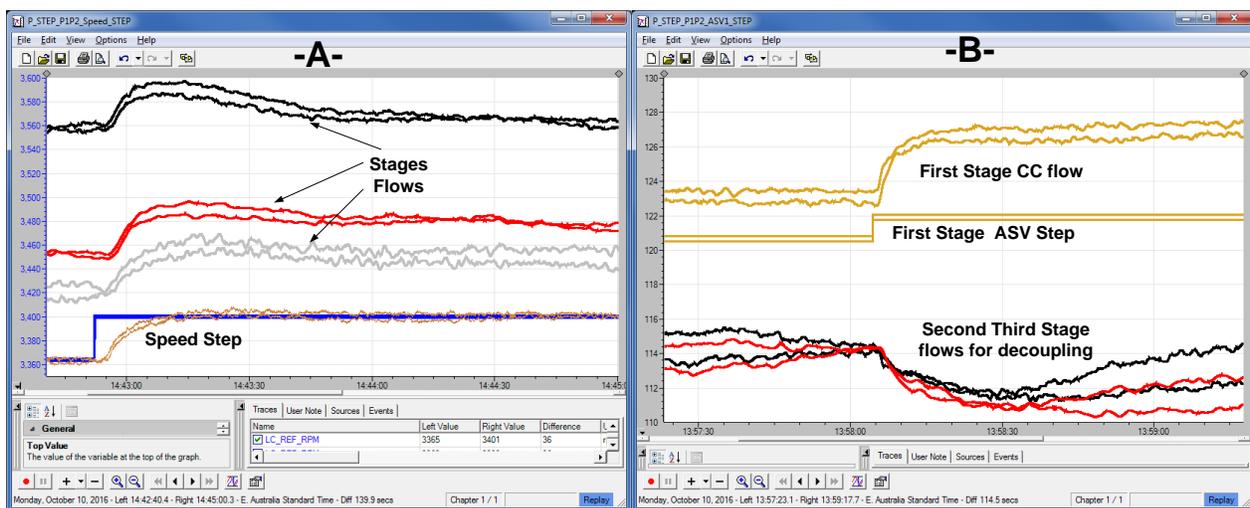
Initial controller settings are retrieved from the simulator testing phase. Commissioning tests, needed to validate controller setting and eventually to update them to final values, are summarized as follows:

-First part of commissioning tests is dedicated to identify and validate the open loop system responses in the real environment by changing controlling variables (i.e. speed, ASV's) with steps. The amplitude of steps is limited (typically 3÷5%) to avoid production disturbances and to provide a local identification. Figure 23 (A, B, C, D) shows typical examples of a recorded open loop responses:

- The speed setpoint from load control is stepped and compressor flow responses are identified to tune speed to ASC decoupling.
- ASV is stepped to identify compressor flow responses of same ASV stage (to tune ASC) and series compression stages (to tune series stages ASC decoupling).
- The speed setpoint from load control is stepped to identify suction controlled pressure response to tune process control.
- ASV is stepped identify suction controlled pressure response to tune process control. Here note also that intervention of quench control to cool compressor suction disturbed the identification. This to underline once more how important is cooperation between operation and commissioning teams.

-Then, the controller's tuning values are to be re-calculated based on the identified process open loop responses.

-After tuning setting recalculation is complete, the system's response to demand changes in closed loop control is checked and validated. The validation test trended in Figure 24 consists in setpoint tracking capability check of the closed loop system when compressors are working in the split region so that both ASV and speed regulators are active. Obtained time constant is checked against design parameter α , that also gives info on disturbance rejection capability of tuned regulator. Also, the same test validates the load control optimization capability to find the operative condition at minimum ASV opening without affecting the suction pressure control.



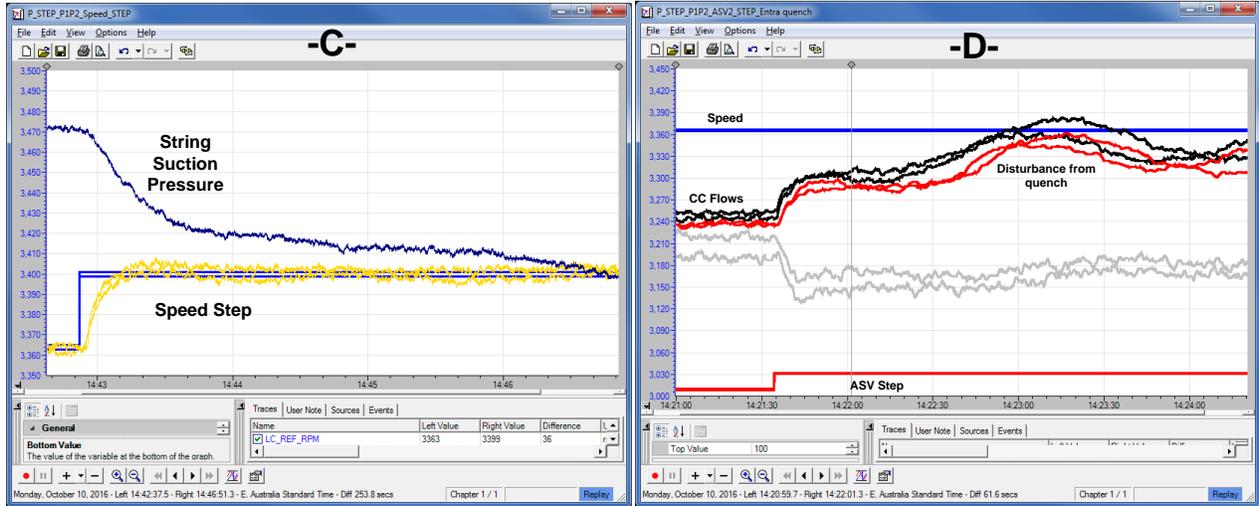


Figure 23 – Step tests (A-B-C-D) for final tuning at commissioning

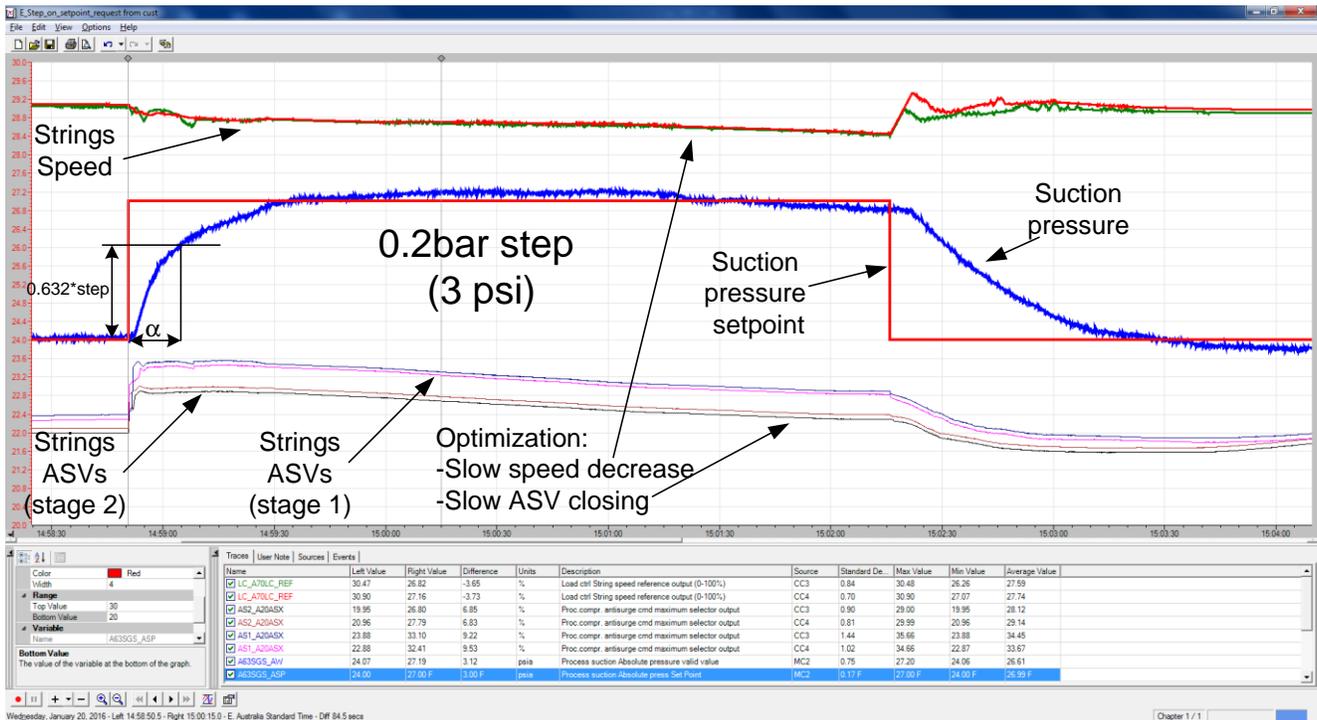


Figure 24 – Suction pressure closed loop validation at commissioning



CONCLUSIONS

Leveraging on knowledge gathered in operating LNG plants, several operability, system availability and reliability improving opportunities were identified and implemented in a new multi-compressor string based grass roots facility. The paper has described the successful execution of a control systems project from the early concept selection phase to front-end and detail engineering, testing, commission, start-up to normal operation.

Compressor aerodynamic design knowledge, understanding of plant process dynamics and careful ASV selection was found to be crucial in modelling the plant and in development of specific control strategies to face critical scenarios. Also, compressor knowledge was key to improve operability of side stream compressor, typically used in LNG propane compressor designs. In-depth understanding of the compressor performance also permitted to design strategies to operate safely and with minimum efficiency impact in case of failure of compressor instrumentation.

Plant controls commissioning procedures were designed to minimize impacts to production. “Trial and error” adjustments are not acceptable in this phase. Instead the plant transient behavior was identified and then the relevant control parameters calculated to match the dynamic requirements. Factory simulation testing, using dry rig and then the plant simulator have proven invaluable in the process. Ultimately, the plant controls commissioning time frame was minimized resulting in a very few compressor trips. Consequent quicker up time to full production helped the LNG plant to achieve its production targets.

NOMENCLATURE

a_s	= Sound Speed at compressor suction	(m/s)
C_v	=Valve Flow Coefficient	(m ³ /s/kPa)
dp_s	=FE differential pressure installed at suction	(kPa)
dp_d	=FE differential pressure installed at discharge	(kPa)
G	= Mass Flow	(kg/s)
K_{FE}	= Flow Element constant	(m ² /1000 ^{0.5})
k_v	= Isentropic exponent in V	(-)
M_w	= Molecular Weight	(g/mol)
N	= Compressor speed of rotation	(rpm)
N_c	= Compressor corrected speed	(rpm)
P_s	= Suction Pressure	(kPa)
P_d	= Discharge Pressure	(kPa)
Q_{vs}	= Volumetric flow at suction	(m ³ /s)
R	= Gas Constant	(kJ/K/kmol)
T_s	= Suction Temperature	(K)
T_d	= Discharge Temperature	(K)
Z_s	= Gas compressibility at suction	(-)

ASV	= Anti Surge Valve
FE	= Flow Element
OTS	= Operator Training System
POC	=Pressure Override Control
POV	= Parallel Override
PV	= Process Value
SCL	=Surge Control Line
SISO	= Single Input Single Output
SIMO	= Single Input Multiple Output
SP	= Controller Set-point



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