DESIGN AND STARTUP OF A UNIQUE SRU ON AN FPSO

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ABSTRACT

While all sulfur plants are different because no set of feed streams or design requirements is the same, most plants fall into a few basic configurations. The differences between them are usually found in the small details of each particular design.

Occasionally, though, plants do not fit the familiar pattern and present unusual challenges. In such cases, traditional design concepts (which are often used without much consideration of the science and engineering that led to them in the first place) may need to be reexamined. The result can be equally unusual plant designs to handle these unusual challenges.

This paper discusses the design of a unique sulfur plant that is part of a gas/oil treating complex installed on an FPSO (Floating Production, Storage and Offloading) vessel moored off the coast of Indonesia. The acid gas is very lean (7-16 mol% H₂S) and contaminated with heavy hydrocarbons (including BTEX), but local environmental regulations and market conditions required high sulfur recovery efficiency (99%) and high sulfur product purity (99.8%). Pretreatment with regenerable activated carbon was required to prevent fouling of the direct oxidation catalyst bed in the Recycle Selectox™ front-end of the sulfur plant, while conventional and sub-dewpoint Claus catalyst beds were required to achieve the high recovery efficiency. The plant design also addressed the unique issues caused by the motion inherent in any floating processing facility, which affected the design of the equipment and instrumentation in a number of ways, some of which are rather subtle.

This facility successfully started up in mid-2017. Some unusual aspects of startup due to the unique design of the plant are discussed, as well as some interesting operational issues that had to be addressed during startup.

Figure 1 – The Karapan Armada Sterling III and the Sulfur Block
Introduction

An international consortium, Husky CNOOC Madura Ltd. (HCML), is developing the Madura Strait Block BD gas reserves to produce sales gas for buyers on the island of Java in Indonesia. This field is located about 40 miles [65 km] east of Surabaya on Java and about 10 miles [16 km] south of Madura Island. The project includes a wellhead platform (WHP) and an offshore spread-moored tanker converted by Bumi Armada Berhad to serve as a Floating, Production, Storage and Offloading (FPSO) vessel (the Karapan Armada Sterling III, or KAS3) with gas/condensate processing facilities and self-sufficient utility systems (Figure 1, preceding page).

The facility is designed to produce 122 MMSCFD [136,000 Nm³/H] of sweet gas and 8,000 BPD [53 m³/H] of condensate. The unmanned WHP with four well slots and two slots for future expansion was set in 182' [55 m] of water, with the FPSO located 410' [125 m] from the WHP. The well fluids are delivered from the WHP to the FPSO via a 10" multiphase flexible riser.

The main process systems on the FPSO are shown in the block flow diagram in Figure 2 below. The incoming wellhead stream is separated into sour gas, condensate, and produced water. The condensate is stabilized and stored in tanks in the hull, and the stabilizer overhead is combined with the sour gas and treated with amine in the Gas Treating Unit (GTU) to remove the acid gas. The treated gas is dehydrated before passing through a hydrocarbon dewpoint control unit, and then exported via an undersea pipeline to the nearby island.

The other part of the facility and the focus of this paper is the Sulfur Block, consisting of a Carbon Bed Adsorption Unit (CBU), Sulfur Recovery Unit (SRU), Sulfur Degassing Unit (SDU), Tailgas Thermal Oxidation Unit (TTO), and Sulfur Storage. These units process the acid gas from the GTU to produce commercial-grade liquid sulfur for offloading and sales.

![Figure 2 – Process Configuration for the FPSO](image-url)
Sulfur Plant Configuration

The Sulfur Block (see Figure 3 below) is designed to process the acid gas produced in the upstream GTU (plus minor amounts from sour water degassers) to produce up to 25 MT/D of commercial-grade liquid sulfur for sales. Indonesian environmental regulations limit the sulfur dioxide (SO₂) concentration in the incinerated vent gas to 2,600 mg/Nm³ (~900 PPMV). To meet this, the Sulfur Block must convert and recover 99% or more of the H₂S contained in its acid gas feed as molten elemental sulfur. (H₂S is the only sulfur species known to be in the gas.)

The GTU is designed to treat sour gas containing 6-7 mol% CO₂ and about 4,000 PPMV H₂S with MDEA, "slipping" 3-5 mol% CO₂ to the treated gas. The resulting acid gas to the Sulfur Block has a very "lean" H₂S concentration (<16 mol%, dry basis). Whereas conventional SRUs partially combust the acid gas in a burner as the first step in the Claus process, acid gases containing less than ~25 mol% H₂S will not burn properly because the heating value of the acid gas is too low to maintain a stable flame. Although an acid gas enrichment unit (AGE) was considered so a conventional SRU could be used, adding a second solvent system was deemed too costly. Instead, the SRU for this project is based on the direct-oxidation Recycle Selectox™ process licensed by UOP, which reacts the acid gas with air across a catalyst bed rather than in a burner. The Selectox™ and Recycle Selectox™ processes can be used to recover sulfur from acid gas streams containing as little as 2 mol% H₂S. In the recycle version, process gas is recycled to the Selectox™ reactor to dilute the H₂S concentration in the reactor feed and moderate the reactor outlet temperature.

Figure 3 – Process Flow Diagram for the Sulfur Block
One consequence of using a direct-oxidation process is that any contaminants in the incoming acid gas will enter the first catalyst bed in the SRU. Heavy hydrocarbons are among the potential contaminants that are harmful to the catalyst used in this process. There was limited information available on the composition of the sour gas to be treated in the GTU, but the acid gas was expected to contain as much as 1,500 PPMV of hydrocarbons heavier than C5, including BTEX (benzene, toluene, ethylbenzene, and xylene). BTEX is particularly troublesome because it tends to be absorbed in amine treating solutions and leave the GTU with the acid gas, then enter the SRU where it is prone to "crack" and foul the catalyst. The CBU uses activated carbon to remove BTEX and other heavier hydrocarbons. It also includes mechanical separation to remove liquids and particulates from the acid gas.

The front-end Selectox™-Claus section of the SRU converts most of the H2S in the acid gas into elemental sulfur that is then condensed and removed, but is limited by reaction equilibrium considerations to sulfur recoveries in the range of 88-94%. Very high (99.8%+) sulfur recoveries can often be achieved by adding an amine-based tailgas cleanup process on the Claus effluent for conventional SRUs, but this will not work with lean acid gas streams like this one because the high CO2 concentration significantly impairs solvent selectivity. Also, tailgas cleanup units are not only expensive to build, but expensive to operate as well and occupy substantial plot space.

A better choice of technology for the intermediate sulfur recovery range of 97-99% is the so-called "sub-dewpoint" Claus process. This process extends the capability of the Claus process by operating the Claus reaction at a lower temperature, so that the sulfur produced by the reaction condenses. Since the Claus reaction occurs in the gas phase, this liquid sulfur does not inhibit the reaction like sulfur vapor does, resulting in a favorable shift in the reaction equilibrium and higher sulfur conversion. The back-end of this SRU is a sub-dewpoint Claus section that uses the CBA (Cold Bed Adsorption) process licensed by BP Amoco to "clean up" the tailgas by recovering additional molten sulfur, raising the overall sulfur recovery efficiency to 99% or higher on a time-weighted average basis.

The SRU contains four catalytic reactor stages: two front-end stages (a Selectox™ stage and a conventional catalytic Claus stage) followed by two back-end catalytic CBA stages. This reactor configuration is similar to the schemes employed by most CBA plants in operation today, except that the first reactor is a Selectox™ stage instead of a conventional Claus stage.

The cyclic operation of CBA catalyst beds is analogous to the adsorption/desorption operation of molecular sieve dehydrators. The CBA reactors are regenerated by using switching valves to divert the Selectox™ effluent gas through a reactor to heat its catalyst bed and vaporize the sulfur to desorb it from the catalyst. The sulfur vapor leaves with the gas and is condensed in the first pass of the sulfur condenser. Part of the gas from the condenser flows to the Claus stage, while the rest is sent back to the Selectox™ stage by a recycle gas blower to dilute the acid gas.

The sulfur produced in the Selectox™ and Claus stages and desorbed from the CBA stages is condensed and removed in the sulfur condenser and flows through sulfur drain traps into the surge tank. This raw sulfur product is pumped to the SDU for degassing, and the degassed sulfur flows into steam-heated storage tanks in the ship's hull. The tailgas from the SRU is routed to the TTO for incineration of the residual H2S and other sulfur compounds. The hot incinerated vent gas is used to generate HP steam for use in the SRU, and then vented to the atmosphere from the top of the vent stack.
Unusual Features in the Sulfur Block

The Sulfur Block for this project is different from conventional sulfur plants in a number of ways. All of these differences were driven by specific project requirements, local environmental regulations, or the service environment (i.e., offshore marine).

Pretreatment of the Acid Gas

In most sulfur plants, all that is required before the acid gas enters the SRU is to remove any liquids that may be present due to carry-over or condensation. While that is also a first step for this SRU, it is also necessary to remove heavier hydrocarbons (especially BTEX) before the acid gas enters the SRU, since the acid gas flows directly to the Selectox™ catalyst bed.

Activated carbon is used in many processes to remove contaminants by adsorption, using intermolecular forces to bind contaminants to the surface of the carbon. There are various techniques for "activating" carbon, but the intent for all of them is to create micropores within the carbon to provide a large amount of surface area to attract and bind the contaminants. In many cases (including hydrocarbon removal), it is possible to alter the operating conditions to desorb the contaminant from the carbon (i.e., regeneration) so that the same carbon can be used multiple times. For such systems, it is typical to arrange the carbon in multiple beds to allow continuous processing, so that a bed can be taken off-line for regeneration while the other bed(s) remain on-line to adsorb the contaminants.

A very small number of SRUs have used activated carbon to remove hydrocarbons from acid gas. The CBU for this project is configured as a 2-bed system to minimize the deck space required. This means there is one bed on-line (adsorbing) and one bed off-line (regenerating) during each cycle. Once the activated carbon in a bed has reached its capacity for removing hydrocarbons, the adsorbed hydrocarbons can be desorbed from the carbon by raising the temperature of the bed. This is accomplished by flowing LP steam (40 PSIG, [2.8 bar(g)]) through the off-line bed. Hot carbon does not adsorb hydrocarbons very well, so the bed is cooled with nitrogen before it is returned to service. The beds are then switched to place the freshly regenerated bed on-line and switch the other bed off-line so it can be regenerated.

Selectox™ Front-End with Recycle Blower

In conventional Claus sulfur plants, the first processing step is a reactor furnace to partially oxidize the acid gas stream. In the furnace, a portion of the inlet H₂S is oxidized to SO₂, which then reacts with the remaining H₂S via the Claus reaction to form elemental sulfur. However, if the H₂S in the acid gas is not at least 25 mol%, a furnace using air (not oxygen) for combustion generally cannot maintain a stable flame and the conventional Claus process cannot be used.

The Selectox™ sulfur recovery technologies are direct-oxidation processes for acid gases with lower H₂S concentrations, as they do not require burning the acid gas first. Instead, a special catalyst (UOP’s proprietary Selectox-33) is used to catalyze both the direct-oxidation of H₂S to sulfur and the conversion of H₂S to sulfur via the Claus reaction:

\[
\begin{align*}
H₂S + 1/2 O₂ & \rightarrow 1/n S_n + H₂O \\
H₂S + 3/2 O₂ & \rightarrow SO₂ + H₂O \\
2 H₂S + SO₂ & \rightarrow 3/n S_n + 2 H₂O
\end{align*}
\]
Since 15-16% H₂S concentration was expected in the acid gas, this SRU uses the Recycle Selectox™ technology. As with any direct-oxidation Claus process, the maximum H₂S concentration in the acid gas that can be processed is limited to about 5% by the exothermic temperature rise across the catalyst bed. The Recycle Selectox™ process overcomes this limitation by diluting the acid gas so that the H₂S concentration of the mixture entering the catalyst bed is about 5%, using some of the partially reacted process gas from downstream of the catalyst bed as the diluent. A steam-jacketed single-stage blower (with installed spare) recycles a portion of the process gas leaving the sulfur condenser pass downstream of the Selectox™ catalyst bed to provide the dilution.

**CBA Back-End Using Selectox™ Effluent for Regeneration**

The front-end section of the SRU includes both a Selectox™ catalyst bed and a conventional Claus catalyst bed. The front-end converts most of the H₂S in the acid gas into elemental sulfur that is then condensed and removed, recovering ~90% of the entering H₂S as molten sulfur. However, Indonesian environmental regulations for the SO₂ concentration in the incinerated vent gas require ~99% recovery for the Sulfur Block to be in compliance.

Adding more conventional Claus reactors to the front-end would not increase the recovery efficiency above 92-94% due to reaction equilibrium. However, the "sub-dewpoint" Claus process can overcome this limitation by operating the Claus reaction at a lower temperature so that the sulfur produced by the reaction condenses. By adding a back-end sub-dewpoint CBA section to recover additional molten sulfur, the overall sulfur recovery efficiency can be increased to 99% or higher on a time-weighted average basis.

A CBA reactor is operated at low temperature (250-300°F [120-150°C]) initially so that it is below the sulfur dewpoint of the reaction products (i.e., "sub-dewpoint") and the sulfur formed is condensed and adsorbed on the catalyst. After operating in this manner for a period of time, the CBA reactor is "regenerated" by routing hot gas through the reactor to vaporize the adsorbed liquid sulfur, which is then condensed and removed in the sulfur condenser. The temperature of the CBA regeneration gas needs to be 550°F to 700°F [290°C to 370°C] to vaporize and desorb the sulfur on the catalyst in a timely fashion. Since the effluent from the Selectox™ catalyst bed is maintained in this range by adjusting the speed of the recycle blower, it is an ideal gas for regenerating the CBA catalyst beds. All that is required is to divert this hot gas through a CBA bed when it needs to be regenerated, and then capture the desorbed sulfur in the sulfur condenser.

**Special Metallurgy**

Conventional Claus plants use carbon steel construction almost exclusively. Certain sections of the plant (generally the furnaces) require special attention because of the high temperatures (1000+°F [540+°C]) involved. This is typically handled by using stainless steel for burner tips and lining the carbon steel furnace shells with alumina refractory (brick usually, sometimes castable). Reactors are usually carbon steel, sometimes with a thin castable refractory lining, and typically have carbon steel or stainless steel grating/screen for holding the catalyst in place, with the structure below the bed supports made of carbon steel.
As long as the sulfur plant refractory linings remain intact to protect the carbon steel from high temperatures, corrosion rates are usually very low, perhaps 1-10 mils (0.025-0.25 mm) per year. This is because the sulfur in the process gases forms a passivating layer of iron sulfide on the exposed surfaces of the carbon steel that is then a barrier to further corrosion. Usually the only time a conventional Claus sulfur plant suffers significant corrosion is when it is not operating, and is allowed to cool down to the point where liquid water condenses and collects. When this happens, the water can form a number of very corrosive acids (including polythionic acid) by combining with the sulfur (and sometimes oxygen) present in the plant, which then rapidly corrode the carbon steel surfaces. This can be prevented, of course, by taking measures to prevent water from condensing and accumulating, such as keeping the plant hot when it is out of service or by purging the plant with inert gas.

Sub-dewpoint Claus plants require more attention to corrosion protection because of the thermal cycling in the sub-dewpoint section. Temperatures in this section are generally in the range of 250-300°F [120-150°C] during the adsorption phase when sulfur accumulates on the catalyst, then up to 550-700°F [290-370°C] during the regeneration phase when the sulfur is desorbed from the catalyst and recovered. As the carbon steel equipment and piping expands and contracts due to the temperature changes, the iron sulfide film tends to crumble and be dislodged, constantly exposing fresh metal to the process gases and leading to high corrosion rates.

In order to avoid rapid corrosion like this in the thermally cyclic sections of the plant, either a corrosion-resistant alloy must be used or the exposed surfaces of the carbon steel must be protected. Large surfaces, like the interior of the reactor vessels, are usually lined with a thin layer of castable refractory. The refractory lining stabilizes the temperature of the steel shell to reduce the amount of expansion and contraction by insulating the shell from the thermally cyclic conditions inside the reactor, so that the iron sulfide layer can remain intact on the steel and protect against further corrosion. The catalyst bed supports are usually 300-series S.S. grating, since the grating components are relatively thin. The reactors in this plant are protected this way.

For piping, one option is to use a diffusion metallizing process to form an aluminum-iron alloy on the surface of the carbon steel for corrosion protection (the Alonizing™ process from Endurance Technologies, for example.) Some companies, though, have found that using corrosion-resistant alloy piping is a more attractive choice because of the logistics and/or cost of this treatment. Either 321 S.S. or 347 S.S. is recommended for this service over other 300 series stainless steels due to their greater resistance to stress corrosion cracking (SCC) induced by polythionic acids. These alloys have been stabilized (via titanium for 321 S.S. and columbium/tantalum for 347 S.S.) to prevent carbide precipitation that can lead to intergranular SCC when exposed to polythionic acid. For this plant, 321 S.S. was selected for the piping.

Some of the equipment and piping in the CBU is also subjected to thermally-cyclic conditions as the beds switch between adsorption and regeneration. Unprotected carbon steel could not maintain a stable iron sulfide film under these conditions and would suffer rapid corrosion, so 316L S.S. is used for the equipment and piping in the thermally-cyclic areas. Since it is unlikely that oxygen will enter this section of the Sulfur Block, the potential for forming polythionic acid is low, so the more expensive 321 S.S. and/or 347 S.S. alloys were not justified. However, the 316L S.S. equipment and piping in the CBU could experience SCC due to the chloride salt present in the marine environment, so all of it was externally coated with Intertherm 228HS to protect against chloride exposure.
Challenges and Solutions for the Sulfur Block

The Sulfur Block also had to address some issues not encountered in more traditional sulfur plants, like those associated with being mounted offshore on a floating vessel or with having uncommon pretreatment requirements. These issues affected the design of the equipment and of the instrumentation mounted on the equipment.

Dealing with Motion

Although an FPSO is anchored in place, it is still subject to motion. Whether it uses a static positioning system like KAS3 or a dynamic positioning system like some production platforms, the ship and everything on it is subjected to both longitudinal (pitch) and rotational (roll) motion. For KAS3, the design basis was a normal range of 1.5° for pitch and 3° for roll, with a maximum range of 4.5° and 9°, respectively.

For process units mounted on ships, the main concern is how this affects liquids in the plant:

- **Liquid Sulfur Rundown Lines**: The rule-of-thumb for sulfur rundown lines (from the condenser to the tank) is to slope them \(\frac{\frac{1}{8}}{\text{" per foot}}\) [\(10 \text{ mm per m}\)] for reliable gravity flow. This suggests the lines should be snapped more on an FPSO so the sulfur can keep flowing when the ship pitches and/or rolls. However, the ship will pitch and/or roll back the other way shortly, so additional slope is not really necessary. The sulfur may pool a bit while the ship moves one way, but the large separator chambers in the sulfur condenser have ample volume, and the sulfur will resume flowing as soon as the ship moves back the other way. As such, the rundown lines on KAS3 do not slope more than in a normal sulfur plant.

- **Sulfur Drain Seals or Sulfur Drain Traps**: Both sulfur drain seals (i.e., U-tubes) and sulfur drain traps (i.e., float-type) have been used in sulfur plants to allow liquid sulfur to flow out of the sulfur condenser without allowing process gas to escape. Drain seals often include a "look box" to verify that sulfur is flowing, whereas drain traps usually have a clear viewport for this purpose. Since the sulfur surge tank for this plant is hung underneath the lower level of the process modules (i.e., above the main deck of the ship), sulfur spilling from a look-box would be a safety hazard, so drain traps are used rather than drain seals.

- **Boiler Levels**: Most boilers in SRUs and TTOs have shutdowns based on the water level in the boiler, measured using instruments mounted on the side of the boiler shell or drum. If the boiler is moving, though, the water level at the side of the shell/drum will also be moving, complicating the measurement.

This movement is addressed in two ways. First, the boilers have horizontal baffles in the water space to dampen the movement and minimize sloshing. The baffles are cut out in the center to allow the steam bubbles being generated in the boiler to escape. Second, the low-low level shutdown points were raised upward so that the low point of the level would be adequate even when the shell/drum pitched or rolled away from horizontal. See Figure 4 on the next page showing how this affects the sulfur condenser tubesheet layout.
In addition, the other alarm points have to be located higher on the shell/drum to prevent nuisance alarms when the ship rolls the other way and the level moves up in the level instrument. Note the larger boiler diameter required to give the same disengaging space above the high level alarm point.

- **Sulfur Tanks**: If the liquid sulfur in the surge tank and the storage tanks is allowed to move freely as the ship pitches and rolls, the sloshing sulfur can develop considerable momentum due to its high density. Not only does the movement complicate level measurement, if the amplitude of the movement is high enough, the liquid momentum could damage the equipment. For instance, the sump-type pumps feeding the SDU that are suspended from nozzles on top of the surge tank could be damaged or dislodged. To prevent this, the tanks have vertical and horizontal baffles to dampen the movement. The baffles have openings in them to allow the sulfur to fill the tanks evenly.

- **Reheat Exchangers**: The heat exchangers for the feeds to the Selectox™ and Claus catalyst beds use HP steam as the heat source. The heat input is regulated by "flooding the bundle", using a control valve to restrict the flow in the condensate outlet line from an exchanger so that a level of condensate accumulates in the exchanger shell. Because this pool of water has less than 1% of the heat transfer rate compared to the steam condensing on the uncovered tubes, the covered tubes have essentially zero heat input.

However, if the ship is pitching and/or rolling, the condensate inside the horizontal exchanger shell will be moving up and down on the tubes, covering and uncovering different tubes and different amounts of tubes. Since the exposed surface area of a horizontal tube does not change linearly as the water level moves up or down (see Figure 5 on the next page), this movement could change the heat input.
To investigate this potential issue, roll effects and pitch effects were considered separately first. All three exchangers are oriented with their longitudinal axes parallel to the length of the ship, so ship roll results in rotation of the water level inside the shell. Although the calculations are rather involved, it is possible to compute the surface coverage for a given volume of water in the shell at different rotations as shown below in Figure 6. It is possible to do the same for pitch (which tilts the shell) as shown in Figure 7 on the next page.

Figure 5 – Covered Area vs. Liquid Height for Tubes

![Figure 5](image)

Clearly, neither roll nor pitch by itself has any significant impact on the amount of tube surface area exposed to live steam. Given the large number of tubes and that some tubes are being uncovered while others are being covered, these results are not surprising. The calculations required to compute the exposed surface are far more complex if roll and pitch are varied simultaneously, but the results were expected to be similar to those for roll and pitch for the same reasons, so no attempt was made to investigate this further.

Figure 6 – Effect of Roll on Surface Area Exposed to Live Steam

![Figure 6](image)
Sulfur Storage and Handling

Offloading the products from an FPSO requires maneuvering another vessel alongside while the product is transferred from one vessel to the other, so weather and sea conditions can delay making the transfer. Since these delays can be extensive (like those due to a hurricane or typhoon, for instance), an FPSO must have ample storage for all of its products so that production is not impacted by delays in offloading. For the sulfur on this FPSO, the design sulfur production rate of 25 MT/D equates to a liquid volume of ~14 m³ (~500 ft³) of sulfur per day, or ~420 m³ (~14,800 ft³) of liquid sulfur in a month. This sulfur must be stored and handled onboard in some fashion before being offloaded by some means.

- **Solid Storage and Handling:** One option is to solidify the sulfur, then store and handle it as solid prills or pastilles. While this may seem to be an attractive option, it is not particularly well suited for offshore facilities due to space constraints and other factors.

  A prilling or pastillation unit is required to form the solid sulfur, which increases the deck space required and adds weight to the FPSO. Solid forming is generally not a continuous operation, so the operators must start up and shut down the unit on a regular basis. After forming, the solid prills or pastilles must be bagged and manually moved (forklifts are not recommended on moving vessels) to the storage area, or conveyed to a storage bin or silo if bulk storage is used. The storage area, which also requires considerable deck space, must be protected from rain and sea spray. Accumulation of sulfur dust is always a concern, as this creates an explosion hazard.

  For offloading to a vessel alongside, bags must be palletized and moved via multiple crane lifts, while bulk solid material must be pneumatically conveyed. The loading area on the other vessel must also be protected from rain and sea spray. Once the transport vessel reaches port, the offloading steps must be repeated to get the sulfur onshore. All in all, a lot of steps and a lot of manual labor on moving vessels are involved in getting the degassed sulfur from the SDU loaded onto a transport vessel and then unloaded in port for this option.
Liquid Storage and Handling: For marine transport, storing and handling sulfur in liquid form has been common using barges since the 1950s and using ships since the 1960s. The first Claus plant on an FPSO, the OS&T (offshore storage and treating) vessel operated by Exxon offshore of California from the early 1980s to the mid-1990s, stored and offloaded its sulfur product in molten form as well. Liquid sulfur is easily transported using heated pumps and piping, and is the most common means of storage and transport for onshore facilities, but it does require having a reliable heating medium.

On the OS&T, transfer from the FPSO to a barge alongside was accomplished using a flexible stainless steel hose. The hose was free-draining from the last block valve on the ship to the vessel so that it emptied completely when loading stopped. A thin layer of sulfur would solidify inside the hose after each loading operation, but it broke up easily as the hose was wound back onto its reel. The sulfur tanks on the barges were heated to keep the sulfur molten, so that it could be pumped onshore when the barge reached port.

Liquid storage and handling was selected for KAS3 because of its simplicity. A surge tank holds the raw sulfur production from the SRU before it is pumped to the SDU. The degassed sulfur from the SDU drains by gravity into either of two rectangular storage tanks in the hull (each sized for 2 months of sulfur production), until it is offloaded using centrifugal pumps. External heating panels using LP steam keep the sulfur molten in the surge tank and the storage tanks.

Carbon Adsorption Design

The carbon vendor for the CBU specified the amount of carbon required and the bed dimensions based on a 24 H adsorption cycle. A 3-bed system was originally proposed, but was changed to a 2-bed system to minimize the deck space needed for the CBU. However, very few details were provided regarding the design of the regeneration system for the carbon.

Fortunately, a considerable amount of design information has been published by Saudi Aramco concerning removal of aromatics from acid gas using activated carbon. This literature gives key design parameters for heating and cooling the carbon, as well as a PFD for the system.

Key Features of the PFD: Figure 8 is an excerpt from Figure 3 showing just the carbon beds and the equipment used during regeneration. Some of the more important features of the design are:

- Acid gas flows upward through the carbon during adsorption
- Steam flows downward so condensate will not accumulate in the vessel
- Nitrogen also flows downward so the top of the bed is as cool as possible when switched back to adsorption
- A condensate drum is underneath the carbon beds for gravity drainage of condensate at the start of heating

![Figure 8 – Process Flow Diagram for CBU](image-url)
The importance of the last point is that when heating first begins, the carbon bed vessel and all its contents are at essentially ambient temperature. The steam condenses immediately on the "cold" steel and carbon, causing the pressure inside the vessel to drop to a vacuum (on the order of 2.2 PSIA [0.15 bar(a)]). Until the vessel and its contents heat up, the vessel remains under a vacuum, so gravity flow is the only way to drain the condensate from the vessel. As the carbon bed heats up, the pressure will increase and eventually rise to the same pressure as the LP steam supply. Until then, the condensate that forms cannot be removed from the condensate drum, so it is sized large enough to hold this condensate until the pressure is high enough to drain the drum on level control.

The purpose of the pressure control drum and its controls may not be immediately obvious. Once the carbon bed comes up to temperature, steam continues to flow through the bed to desorb the hydrocarbons from the carbon. During this time, little condensation of the steam occurs in the vessel, so it is condensed in the exchanger downstream of the condensate drum. Since this would likewise cause a vacuum in the pressure control drum and make draining the condensate difficult, nitrogen is added as needed to maintain positive pressure downstream of the condenser. However, when the carbon bed is later cooled by flowing nitrogen through it, no condensation occurs and the nitrogen must be vented to the flare instead through the other pressure control valve to maintain pressure.

- **Unsteady-State Operation:** Heating and cooling of the carbon beds during regeneration are inherently unsteady-state processes. Estimating the rates at which these occur is important for two reasons. First, heating and cooling of the off-line bed must be accomplished within the same time frame as the adsorption phase of the on-line bed, so the rates must be sufficient for the regeneration cycle time not to exceed the adsorption cycle time. Second, these rates are a function of the flow rates of the steam and nitrogen, and these flow rates must be compatible with the utilities system limitations of the FPSO.

During the unsteady-state heating phase, the steel, carbon, and support media of the carbon bed (plus the steel in the condensate drum) start at ambient temperature and heat up as the steam is condensed. The temperature rise during a given time interval depends on how much latent heat is released as the steam condenses, which is a function of the condensing temperature during that time interval. Since the condensing temperature is changing continuously, the heating rate is changing continuously, which complicates the calculations.

One way to simplify the calculations is to break the heating period into short time intervals so that the condensing temperature does not change much during each interval. With this approach, the calculations indicated that the bed temperature would reach steady state in less than one hour based on the steam flow rate set using the steam:carbon ratio suggested by the Saudi Aramco data. This was considerably less than the heating cycle time that had been assumed, meaning that the carbon would be at its maximum temperature during most of the cycle, which should provide ample time for the adsorbed hydrocarbons to desorb.

A similar approach was used for estimating the cooling phase. Unlike heating with steam, though, cooling with nitrogen does not utilize latent heat, only sensible heat, so cooling proceeds at a much slower rate. The nitrogen flow rate was adjusted until the bed
cooled down to about the same temperature as the acid gas within the allotted cooling time (16 H). Unfortunately, this nitrogen flow rate exceeded the capacity of the nitrogen generators selected for the FPSO utility systems. When the flow rate was reduced to what was available from the nitrogen generators, the simplified calculations indicated the carbon bed could not be cooled completely in the time allotted for the cooling cycle.

A more complex calculation model was developed that also broke the carbon bed into small intervals (slices) so that the temperature change of each slice was computed during each time interval, with the resulting nitrogen temperature then used for the next slice. Not only should this improve the accuracy of the prediction, it would also allow observing the temperature behavior at different heights inside the bed. As seen below in Figure 9, the more complex model also indicated that the bed could not be cooled completely.

The model did show, however, that the top half of the bed would be cool when the freshly regenerated bed switched on-line. At that point, acid gas would begin flowing upward through the bed and continue cooling the carbon. Because the acid gas mass flow rate is much higher than the nitrogen flow rate, the cooling rate is much faster, and the entire bed would be cool within about 2 H. Although a heat "front" does make its way upward through the bed, it was hoped that its magnitude would not impair BTEX removal. In hindsight, deviating from the vendor's recommendation and having the nitrogen flow the same direction as the acid gas (upward) would have been better.

Figure 9 – Estimated Cooling of the Carbon Bed
Plant Startup and "Lessons Learned"

Preparations for startup of the Sulfur Block began in May 2017 in parallel with startup of the other process systems. Operational issues in some of the other process units took time to resolve, so acid gas introduction was delayed until late June 2017. Startup proceeded smoothly, and a successful performance test was completed in mid-July 2017. Due to onshore limitations on treated gas takeaway, the sulfur production rate was ~20% of design during most of this period.

Certain aspects of startup for the Sulfur Block are a bit different from a conventional sulfur plant, two of which are discussed below. There were also some aspects of the plant design that did not perform as expected and required some adjustments, also discussed below.

Startup Sequencing

In most sour gas plants, the Sulfur Block is warmed up so that it is ready for acid gas before introducing sour gas to the GTU. For a "black" start on the FPSO, however, there is no fuel gas because it comes from the treated gas leaving the GTU. This required "boot strapping" the startup of the GTU, SRU, and TTO for the initial startup.

The turbine generators and the auxiliary boiler on the FPSO are designed to operate on either fuel gas or diesel, so power and LP steam can be generated using diesel fuel to allow circulating and stripping the solvent in the GTU. Once sour gas is introduced into the GTU, treated gas is available for use as fuel gas and the TTO can begin warming up. As the TTO comes up to operating temperature, its waste heat boiler will start producing HP steam, and the SRU can then start warming up by heating nitrogen circulated through the unit with the recycle gas blower. Acid gas can be introduced into the SRU only after warmup of the SRU is complete. Until that time, the acid gas from the GTU must be sent to the flare.

"Long Loop" versus "Short Loop" Recycle in the SRU

Since this SRU does not have a burner, it cannot fire fuel gas to heat up its equipment for startup, so it needs another means of heating the equipment. During normal operation, the recycle gas blower circulates process gas from the first pass of the condenser back to the Selectox™ pass of the reactor, which is termed the "short loop". For warmup, the recycle gas blower can be used to circulate nitrogen through the short loop, which includes the recycle gas heater. HP steam from the TTO will heat the nitrogen as it flows through this exchanger, which in turn heats the catalyst in the Selectox™ pass of the reactor and the water in the sulfur condenser.

For plant warmup, though, circulating through and heating all of the catalyst beds and condenser passes would be better. To accomplish this, the SRU has switching valves to allow the blower to take its suction downstream of the final CBA reactor, the "long loop". However, the longer path meant more pressure drop, and the vendor could not supply a blower that would operate stably under both conditions. The option of using the "long loop" was retained, but an alternate startup scheme was developed in case the selected blower could operate only in the "short loop". Since the "short loop" heats just the Selectox™ pass of the reactor, "cold bed" startup would be used for the Claus pass of the reactor and for the two CBA reactors. During the precommissioning phase of startup, however, the blower was operated in both the "short loop" and the "long loop" without any issues, so the "long" loop has been used for all startups.
"Lessons Learned" for the CBU

Although the CBU generally operated as intended, there were some interesting issues addressed during commissioning and startup.

- **Heating and Cooling Rates:** The actual heating rate for the carbon beds proved to be close to that estimated during the design phase. However, even though there was frequently more nitrogen available for cooling than was assumed for the design, it still was insufficient to fully cool the carbon beds during the 16 H initially set for the cooling cycle. Since the temperature during heating reached steady-state in about an hour, the heating cycle was shortened from 8 H to 2 H so the cooling cycle could be extended to ~22 H. However, the additional time had only a minor impact on the bed temperature, as cooling or heating quickly becomes asymptotic when only sensible heat is available.

What is more interesting is the measured cooling rate compared to the model predictions. Figure 10 shows the predicted temperatures for the top and bottom of the bed (adjusted for the higher nitrogen flow rate) overlaid with the measured gas outlet temperature for the carbon bed. Unlike the predicted bottom bed temperatures (which should be the same as the gas outlet temperature), the measured gas outlet temperature shows an immediate and fairly rapid drop. This quickly becomes a much more gradual rate of decline, though, as the temperature asymptotically approaches the nitrogen supply temperature. Attempts to reconcile the model with the measured temperatures were unsuccessful.

![Figure 10 – Predicted vs. Measured Cooling of the Carbon Bed](image-url)
**Pressure-Equalization Step:** During cooling of the carbon bed, the nitrogen pressure was maintained at 40 PSIG [2.8 bar(g)]. When cooling was complete, the CBU Valve Sequencer (CVS) switched the vessel to "standby" by closing its nitrogen valves. Its acid gas inlet valve was then opened to allow the pressure in the vessel to equalize with the acid gas pressure (0-3 PSIG [0.0-0.2 bar(g)]. (The acid gas outlet valve remained closed, so acid gas would not enter the vessel while it is in "standby".)

However, the vessel de-pressured rapidly when the acid gas inlet valve opened, sending a large flow of nitrogen through the SRU. The resulting increase in the flow measured by the acid gas flow meter caused the air:acid gas ratio control system to add more process air, when in fact the actual amount of acid gas had not changed. The net effect was a large disturbance to the SRU from sending too much air to the Selectox™ catalyst bed.

Slowing down the switching valves was tried first to address this issue, but this made no appreciable difference because these are large ball valves. Instead, the CVS was modified to add a pressure-equalization step prior to switching the regenerated carbon bed to "standby", by venting the nitrogen in the vessel to the flare via the condensate and the pressure control drums. The carbon bed continues to vent until its pressure is the same as the acid gas pressure, then the vent valves are closed and the acid gas inlet valve is opened to put the bed in "standby". These changes completely eliminated this cause of disturbances in the SRU.

**Bed Switching Step:** When one carbon bed switched on-line and the other carbon bed then switched off-line, there was a sudden large drop in the acid gas flow rate that lasted for 1-2 minutes. It appeared that the freshly regenerated carbon has an affinity to adsorb CO₂ and/or H₂S. As it did so, the acid gas flow rate to the SRU dropped almost to zero, which again upset the air:acid gas ratio control for a period. Once the carbon became saturated with acid gas, the flow rate to the SRU returned to its previous value.

Manually stepping through the bed switches showed that pausing after the regenerated bed was first switched on-line (while it was still in parallel with the current on-line bed) gave a much more gradual drop in the acid gas flow rate. This is probably because only a portion of the acid gas flows through the regenerated bed while the two beds are in parallel, so less adsorption of acid gas occurred. During this testing, it was found that the decrease and subsequent increase in the acid gas flow rate occurred slowly enough for the air:acid gas ratio control system to respond and keep the SRU close to on-ratio operation.

The CVS was modified to include a short pause before completing the switch. Although the bed switch still causes a disturbance in the SRU due to the brief drop in acid gas flow rate, the magnitude of the disturbance is much less than before. As shown in Figure 11 on the next page, the flow rate drops by about 40% for a brief period, but then returns to its previous value. (Prior to adding the pause, the drop was almost 100%.)

This figure also shows the effect of the incomplete cooling of the carbon bed: the sudden jump in the acid gas outlet temperature from the CBU when the beds switch, which takes a couple of hours to dissipate. (Note the small temperature dip when the beds first switch, which has been attributed to adsorption of acid gas onto the carbon. After this brief dip, the temperature jumps up by ~10°C due to the residual heat in the carbon bed.)
Lessons Learned for the SRU

Startup of the SRU was quite smooth, but there were some issues that disturbed its operation and affected the sulfur recovery efficiency. Modifications to the control schemes alleviated most of these disturbances.

Although the GTU was designed to "slip" 3-5 mol% CO₂ to the treated gas, the CO₂ removal was much higher during startup with the GTU operating at reduced rates. As a result, the H₂S concentration in the acid gas was much lower during startup (~7 mol%) than the original design basis (~15 mol%). The sulfur production rate was only ~20% of design due to the low GTU throughput, but the process gas flow rate in the SRU was ~40% of design due to the higher CO₂ concentration in the acid gas.

- **Acid Gas Flow Rate Fluctuation:** The GTU was treating the sour gas down to about 0.75-1.00 mol% CO₂ most of the time. Periodically, though, the CO₂ concentration in the treated gas would drop very quickly down to about 0.30-0.50 mol% and remain there for a period, then just as quickly go back up to the previous value. This additional CO₂ removal caused the acid gas flow rate to the SRU to increase when this CO₂ was stripped from the amine. (These fluctuations in the GTU behavior have been attributed to operation at reduced gas flow rates and its impact on selective removal of H₂S.)

However, although the acid gas flow rate to the SRU increased when more CO₂ was removed, the quantity of H₂S in the acid gas was unchanged (meaning the acid gas was leaner during this period). When the feed-forward portion of the air:acid gas ratio control loop saw the increase in acid gas flow, it added more air. Unfortunately, this is not the proper control response, because the quantity of H₂S in the acid gas had not changed. The result was wide swings in the air demand in the SRU, making good control of the
H₂S:SO₂ ratio in the SRU difficult. Due to the inherent time lag in the response of the air demand analyzer, the GTU had gone back to its usual CO₂ removal by the time the analyzer sensed the change in the H₂S:SO₂ ratio. Thus, the feedback control from the air demand controller was too slow to help.

A feed-forward control based on the amount of H₂S that is in the acid gas is needed to correct this issue. The best way to provide this would be to add an H₂S analyzer on the acid gas to give the H₂S concentration in near real-time, which could then be multiplied by the acid gas flow rate to give the H₂S flow rate. Because the O₂:H₂S ratio required by reaction stoichiometry is constant (0.5:1), the required air:H₂S ratio is also constant and the setpoint for the air flow controller could be computed directly.

Since it would take a while to purchase and install an H₂S analyzer, this solution was of no immediate help. As a temporary solution, an algorithm was added in the DCS to estimate the H₂S concentration in the acid gas using the CO₂ and H₂S concentrations measured for the treated gas leaving the GTU. Once this estimated H₂S concentration was added to the feed-forward portion of the air:acid gas ratio control, the fluctuations in air demand due to acid gas flow rate were almost completely eliminated.

- **Selectox™ Reactor Temperature Control:** As discussed previously, the recycle gas blower speed is adjusted to control the outlet temperature from the Selectox™ (1st) pass of the reactor by changing the recycle gas flow rate. However, the response time for the outlet temperature proved to be quite slow. This is due in part to the SRU operating at low throughput, but it is also because the catalyst, steel, and refractory in this chamber of the reactor are a large "thermal mass" compared to the process gas flowing through the catalyst bed. As a result, it takes a long time for process changes that affect the ΔT across the catalyst bed to change the outlet temperature, as the heat "front" has to work its way down through the bed before this temperature responds. It often took as long as 30 minutes for the outlet temperature to respond to a process change, so adjusting the speed of the recycle gas blower to control the outlet temperature did not work very well with the acid gas flow rate fluctuating as it was.

However, the top bed temperatures responded to process changes very quickly, and seemed to be better suited for control purposes, especially at low operating rates. A second temperature controller was configured in the DCS to use these temperatures as an alternative for adjusting the blower speed. For reliability, the controller uses the "middle-of-3" of the three top bed temperatures so that failure of a single temperature element will not hamper control. The new controller worked significantly better, as can be seen in Figure 12 on the next page. (Note the disturbances described earlier due to the CBU bed switches that can also be seen in the figure.)

Most of the variation that remains in the reactor temperatures is likely due to variations in the acid gas flow rate, some of which are due to the variation in the CO₂ pickup in the GTU. Some time was spent trying to develop a feed-forward algorithm for adjusting the blower speed based on the acid gas flow rate and concentration, but with no success. Although the required recycle gas flow rate can be easily computed, there is no means for correlating this flow rate to the corresponding blower speed because there are too many other variables that also affect the flow-head characteristics of the blower.
High SO₂ Emissions after a CBA Bed Switch: The timers in the CBA sequencer were initially set at 12 H for heating and 4 H for cooling. Because of the low acid gas flow, though, the rate at which a CBA reactor cooled was too slow to adequately cool the catalyst bed before the reactors switched. This caused a sharp increase in the SO₂ emissions from the TTO after each bed switch, as seen in Figure 13 below.
From examining the catalyst bed temperatures, it was clear that the entire catalyst bed had been regenerated within 8 H of the beginning of the heating cycle, but the bed had only cooled to 320-390°F [160-200°C] by the end of the 4 H cooling cycle. When the hot bed was then switched to the "lag" position, it could not function as a sub-dewpoint bed until all the catalyst was much cooler, which took several more hours. Until the bed was cool, the SO₂ emissions remained high.

The heating time was shortened to 8 H and the cooling time was extended to 8 H, which kept the overall cycle time the same. This allowed the bed to be much cooler when the CBA reactors switched. Notice in Figure 14 below that the SO₂ emissions went down when the CBA reactors switched, which is what should happen. (The jump in the SO₂ just prior to the CBA bed switch was caused by the disturbance from the CBU beds switching and causing a brief drop in the acid gas flow rate as discussed earlier.)

Subsequent Operations

First sulfur production was in late June 2017 (see Figure 15 below). A 72 H performance test was conducted in July 2017 with the GTU operating at about 50% of design throughput. The acid gas to the SRU averaged about 3.9 MMSCFD [4,400 Nm³/H] and 7.5 mol% H₂S (wet basis) during the test (see Figure 16 on the next page). This acid gas flow rate was about 82% of design, while the corresponding sulfur production (~11 MT/D) was about 45% of design due to the H₂S concentration being about half of the design value.

Plant conditions were very stable during the test run, and the performance of the Sulfur Block exceeded the guarantee requirements. Sulfur recovery efficiency was significantly higher than the required 99.0%. The sulfur product was analyzed at 99.97 wt% purity with less than 5 PPMW of H₂S (versus the required 99.8 wt% and 10 PPMW, respectively).
Routine operation and sulfur production are on-going under the direction of the FPSO operating staff. The first sulfur offloading was successfully conducted in November 2017. Over 300 MT of molten sulfur was transferred from the hull storage tank into a heated barge alongside the FPSO in one hour.

References