Sulphur Plants Produce Valuable Energy – Use it Wisely

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ABSTRACT

Energy conservation has always been important to the oil and gas industry but is becoming increasingly critical in the current low oil price climate. Sulphur recovery facilities are necessary to meet emissions regulations and are therefore often viewed as a cost of production. However, the sulphur plant is normally a net energy exporter, providing a frequently overlooked benefit to the energy balance of the processing complex. This is because the Claus reaction, which is employed to convert H_2S to elemental sulphur, is exothermic, and the waste heat from the process can be recovered as HP and LP steam.

This paper describes the energy benefits of the Claus sulphur recovery process and the net impact that various tail gas treating technologies can have in eroding or enhancing these benefits; this work was previously presented in another paper.[6] In addition to the previous work with conventional tail gas treating technologies, a non-traditional tail gas treating technology is explored in this paper, as a means of improving the energy balance of the overall facility. Benchmark key performance indicators (KPIs) are provided for all technologies, for operator guidance on how their sulphur recovery facilities should be performing, from an energy efficiency perspective.

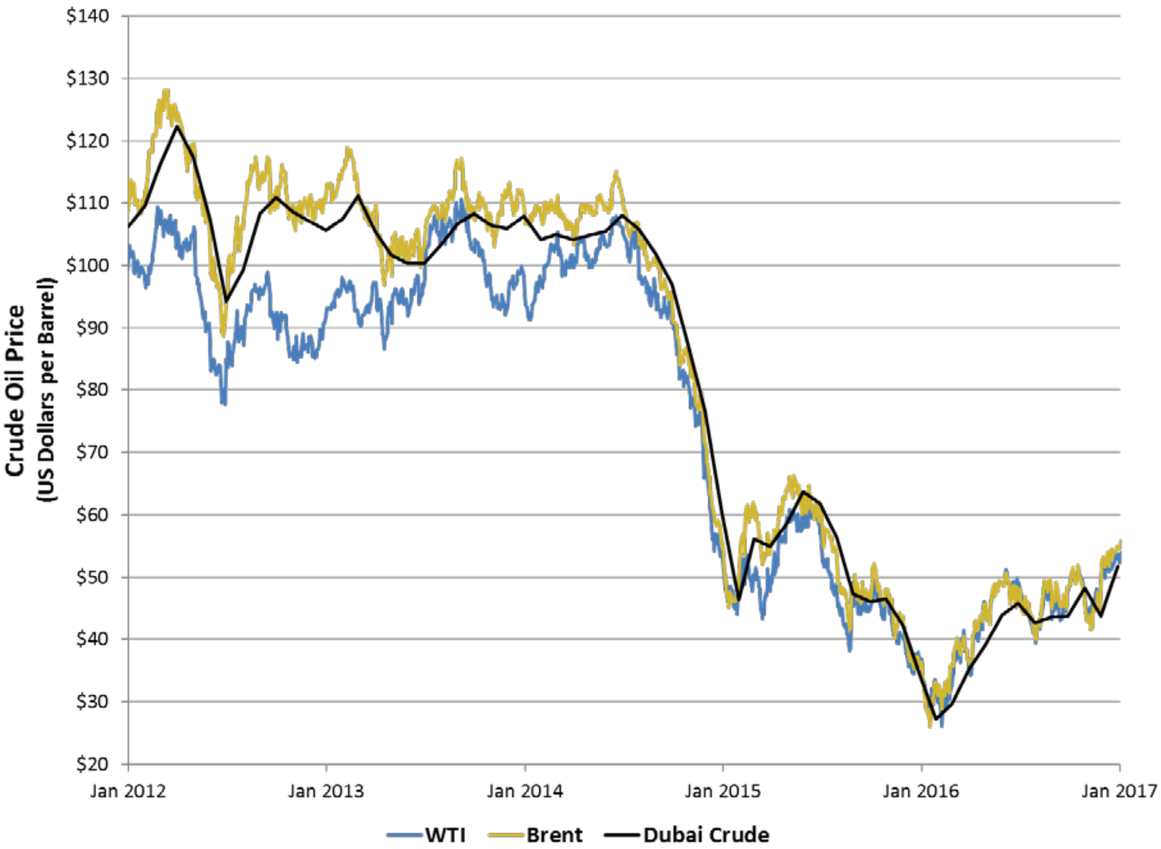
Based on the benchmark information provided in this paper, total SRU energy production for one of the world's largest sulphur producing nations is estimated. Potential future SRU energy production for various tail gas treating technology scenarios is also contemplated.

INTRODUCTION

In late 2014, oil prices plummeted dramatically in the space of just a few months. As shown in **Figure 1**, around mid-year, the price of oil started dropping rapidly and fell from about $100 per barrel to approximately half of this figure in less than six months. The price suffered additional losses in 2015, falling to a minimum around $25 per barrel, the lowest value in more than 13 years. The price rebounded somewhat in 2016 and now sits at around $55 per barrel (January 2017); however, this is still roughly half the figure enjoyed by the industry during the preceding 5 years.

Low oil price creates both challenges and opportunities for the industry. The challenges are obvious, with threatened economics for new projects and reduced margins on existing production. Those producers who stay focused on achieving success via efficient, cost-effective operations during these challenging times will prosper and thrive when oil price rebounds, allowing for stronger and more profitable operations in the future.

Figure 1. 5-Year Historical Crude Oil Price[1,2]



The sulphur recovery facility within a refinery or gas plant is required to meet SO_2 emissions regulations and is often viewed as a cost of production. However, waste heat from the exothermic Claus reaction is recovered as HP and LP steam, which usually makes the sulphur plant a net energy exporter, supplying needed steam and/or power to the processing complex. For extremely sour gas plants or refineries processing sour crude feedstock, the sulphur plant may be one of the areas of greatest interest for improving energy efficiency and strengthening the economics of production.

As such, this paper will:

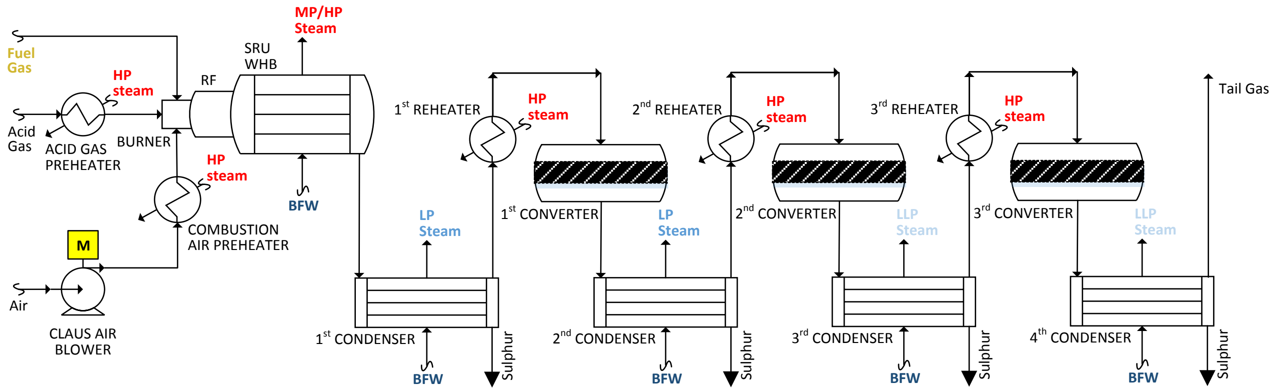
- Identify the top energy producers and consumers in a typical sulphur recovery facility
- Quantify the energy balance and provide benchmark energy performance indicators (KPIs) for a hypothetical Benchmark Plant, across a range of conventional sulphur recovery technologies
- Consider an alternate tail gas treating technology which may improve the energy balance of the sulphur recovery facility, while still achieving high sulphur recovery efficiency (SRE)

This information will be used to examine the current and potential future energy export from all sulphur recovery units (SRUs) in one of the world's largest sulphur-producing nations.

ENERGY PRODUCTION & CONSUMPTION IN THE SULPHUR PLANT

The Modified Claus process is shown in **Figure 2**. In this well-known process, one third of the H_2S in the acid gas is burned to form SO_2, which then reacts with remaining H_2S to form elemental sulphur, via the exothermic Claus reaction. Key utilities produced/consumed in the process are steam (HP and LP), fuel gas and electrical power. Typically, some form of tail gas treating is required downstream of the SRU to satisfy SO_2 emissions regulations, the energy requirements of which can be substantial and will thus be discussed throughout this paper.

Figure 2. PFD of 3-stage Claus SRU with Key Utility Streams



Steam

The heat of reaction for the exothermic Claus reaction is as indicated in Equations 1-3. [3]

Thermal Section

$H_2S + 3/2O_2 \rightarrow SO_2 + H_2O$ $\Delta H@77°F = -223{,}100$ BTU (1)

Thermal and Catalytic Reaction Section

$2H_2S + SO_2 \rightarrow 3/xS_x + 2H_2O$ $\Delta H@77°F = -41{,}300$ BTU (2)

Overall Reaction

$3H_2S + 3/2O_2 \rightarrow 3/xS_x + 3H_2O$ $\Delta H@77°F = -264{,}400$ BTU (3)

As shown in **Figure 2**, heat released in the process is recovered in the SRU waste heat boiler (WHB) as HP or MP steam, and in the sulphur condensers as LP or LLP steam. In addition to heat released by the Claus reaction, the incinerator produces heat via combustion of fuel gas to achieve temperatures hot enough to ensure complete oxidation of H_2S in the tail gas stream. Incinerators are often equipped with waste heat boilers and/or HP steam superheaters to recover some of this heat and maximize process efficiency.

Steam consumers in the process include feed gas preheaters and process gas reheaters, all of which typically utilize HP saturated steam. The SRU and incinerator air blowers may also consume HP steam if steam turbine drives are employed. The only continuous LP steam consumer is the reboiler in an amine-based tail gas treatment unit (TGTU).

Overall, the sulphur recovery facility is a net HP steam exporter. It is typically also an LP steam exporter; however, this may not be the case for amine-based TGTUs with extremely high recovery efficiency requirements, which can consume all (or more) of the LP steam produced in the SRU. This is usually not the case unless SRE significantly exceeds 99.9%.

Fuel Gas

The incinerator is a continuous fuel gas consumer. Fuel gas is burned with excess air and the combustion effluent is mixed with SRU tail gas to achieve a minimum temperature of 650 ^oC for nearly complete oxidation of H_2S to SO_2. Sometimes higher temperatures are required to achieve lower limits on CO and/or total reduced sulphur (TRS), up to a maximum of around 815 ^oC.

In some facilities which process lean acid gas, continuous fuel gas co-firing may be employed in the SRU burner to achieve temperatures high enough for BTEX destruction. Other methods for increasing furnace temperature such as acid gas enrichment (AGE) or oxygen enrichment are preferred, as they reduce the risk of soot deposition and/or fire in the downstream catalyst beds, as well as minimizing the process gas flow through the facility, thereby minimizing the size of equipment and piping. Nevertheless, fuel co-firing is not an uncommon practice for increasing furnace temperature.

In older facilities, fuel gas is sometimes consumed in SRU fired reheaters; however, most modern SRUs utilize indirect HP steam reheaters to avoid the concerns mentioned above for fuel co-firing in the SRU burner. Most modern amine-based TGTUs employ preheating with HP saturated steam upstream of the hydrogenation reactor. However, for facilities that are not equipped with a hydrogen source, reducing gas generators (RGGs) are often installed. In an RGG, fuel is combusted sub-stoichiometrically to produce reducing gas; the exhaust gas is then mixed with the SRU tail gas to achieve sufficient temperature for the hydrogenation and hydrolysis reactions to occur in the downstream reactor. RGGs result in increased energy consumption (vs. TGTU steam preheaters) due to fuel consumption in the burner and also result in increased process gas flow through all equipment downstream of the RGG.

Overall, the sulphur recovery facility is a net fuel gas importer. All SRUs require continuous fuel firing in the incinerator; however, facilities which employ continuous fuel firing in the SRU burner, reheaters and/or TGTU RGG may require significantly more fuel consumption than units which do not.

Electric Power

The Claus and incinerator blowers are the primary electric power consumers in a sulphur recovery facility, when these machines are equipped with motor drivers. Other power consumers include air-cooled heat exchangers and pumps. In hot climates and/or when extremely high sulphur recovery efficiency is required, refrigeration may be required for solvent and quench water cooling in the TGTU.

Overall, the sulphur recovery facility is a net power importer. Facilities which employ amine-based tail gas treating may utilize significantly more power than those which do not, due to additional air cooled exchangers, pumps and possible refrigeration utilized in those facilities.

ENERGY PERFORMANCE COMPARISON OF SULPHUR RECOVERY TECHNOLOGIES

The overall impact of the various utility producers and consumers described above is that the sulphur recovery facility is typically a net energy exporter, although the quantity of energy exported can vary greatly depending on the type of tail gas treating technology employed. In some cases, the facility may actually need to import energy, when very high sulphur recovery efficiency is required, negating the energy benefits of the Claus process. To illustrate this, a hypothetical 1,000 MTPD sulphur recovery train has been considered, over a range of sulphur recovery efficiencies, which will be referred to as the Benchmark Plant. Considering that most refineries produce rich acid gas (H_2S > 85 mol%) and most gas plants produce relatively lean acid gas (40-50 mol% H_2S), a median concentration of 60 mol% is assumed. Feed gas flow and composition for the hypothetical plant are provided in **Table 1**.

Table 1. Feedstock for Hypothetical 1,000 MTPD Sulphur Recovery Train (Benchmark Plant)

	mol%	kmol/h
Component		
H_2S	60%	1,300
CO_2	30%	650
Hydrocarbon (as C_1)	1%	22
H_2O	9%	195
Total	100%	2,167
Temperature, °C	54	
Pressure, barg	0.69	

To compare relative energy balances for varying recovery efficiencies, simulations were generated, over a range of tail gas treating technologies and plant configurations. The following SRE cases were explored.

A. **97% SRE** – 97% recovery is based on a conventional 3-stage Claus unit.
B. **99.0% SRE** – 99.0% recovery is based on a sub-dewpoint process (2-stage Claus + 2 sub-dewpoint reactors), although it should be noted that a direct oxidation process would achieve similar SRE and energy balance.
C. **99.3% SRE** – 99.3% recovery is based on a 2-stage Claus unit + TGTU (MDEA). This SRE is just beyond the upper limit of an achievable guarantee value for sub-dewpoint and direct oxidation processes; therefore, it was investigated as the entry point for an amine-based TGTU.
D. **99.9% SRE** – 99.9% recovery is based on a 2-stage Claus unit + TGTU (MDEA).
E. **150 mg SO_2/Nm^3 (MDEA)** – The World Bank Standard case (99.98% SRE) is first investigated based on a 2-stage Claus unit + TGTU with generic solvent (MDEA).
F. **150 mg SO_2/Nm^3 (Proprietary Solvent)** – The World Bank Standard case (99.98% SRE) is investigated utilizing a more selective solvent in the TGTU and corresponding positive energy impact; thus, this case is based on a 2-stage Claus unit + TGTU with proprietary solvent.

Process flow diagrams for the six cases are provided in **Figures 3-6**.

Figure 3. PFD for Benchmark Plant Case A

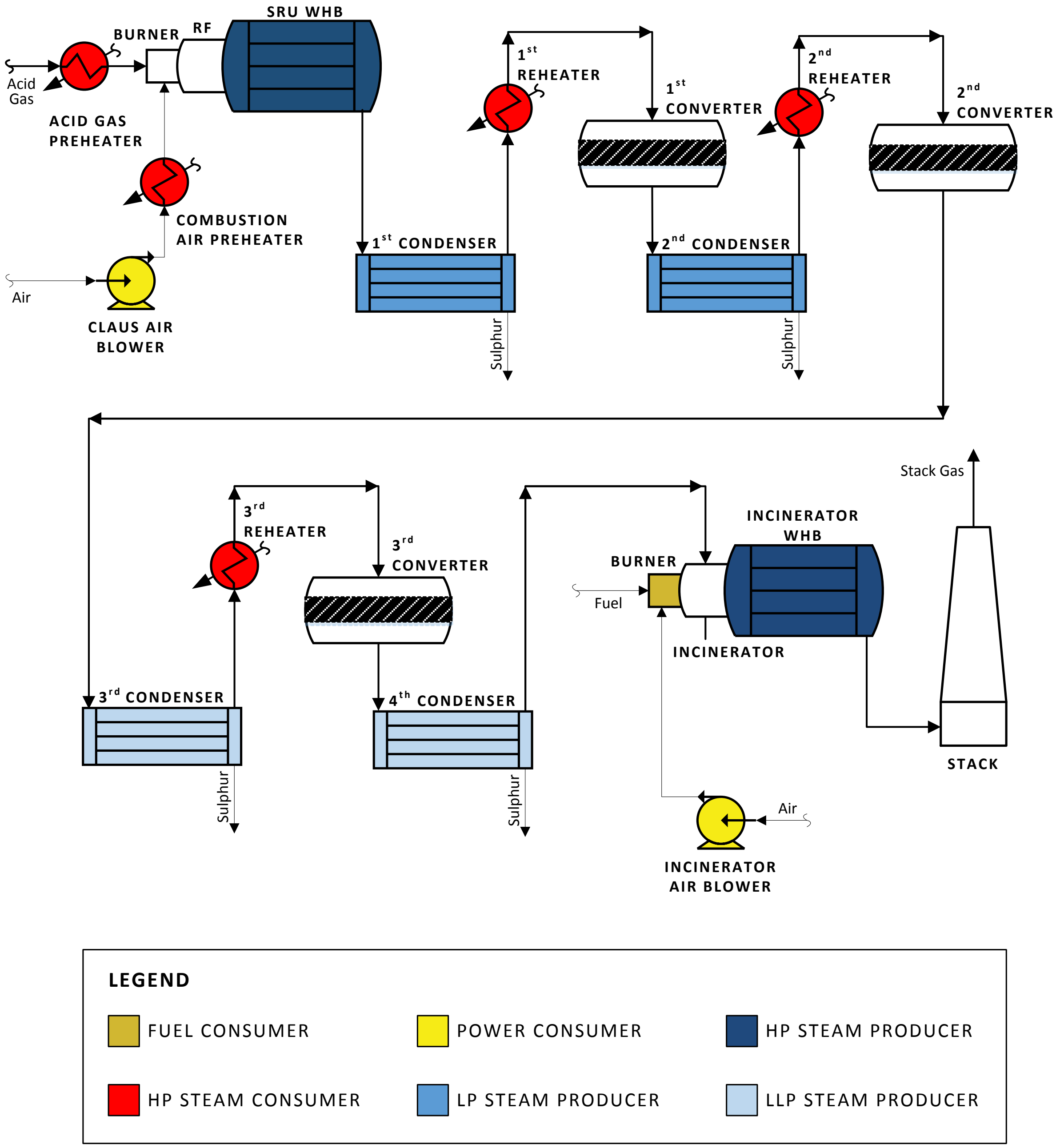


Figure 4. PFD for Benchmark Plant Case B

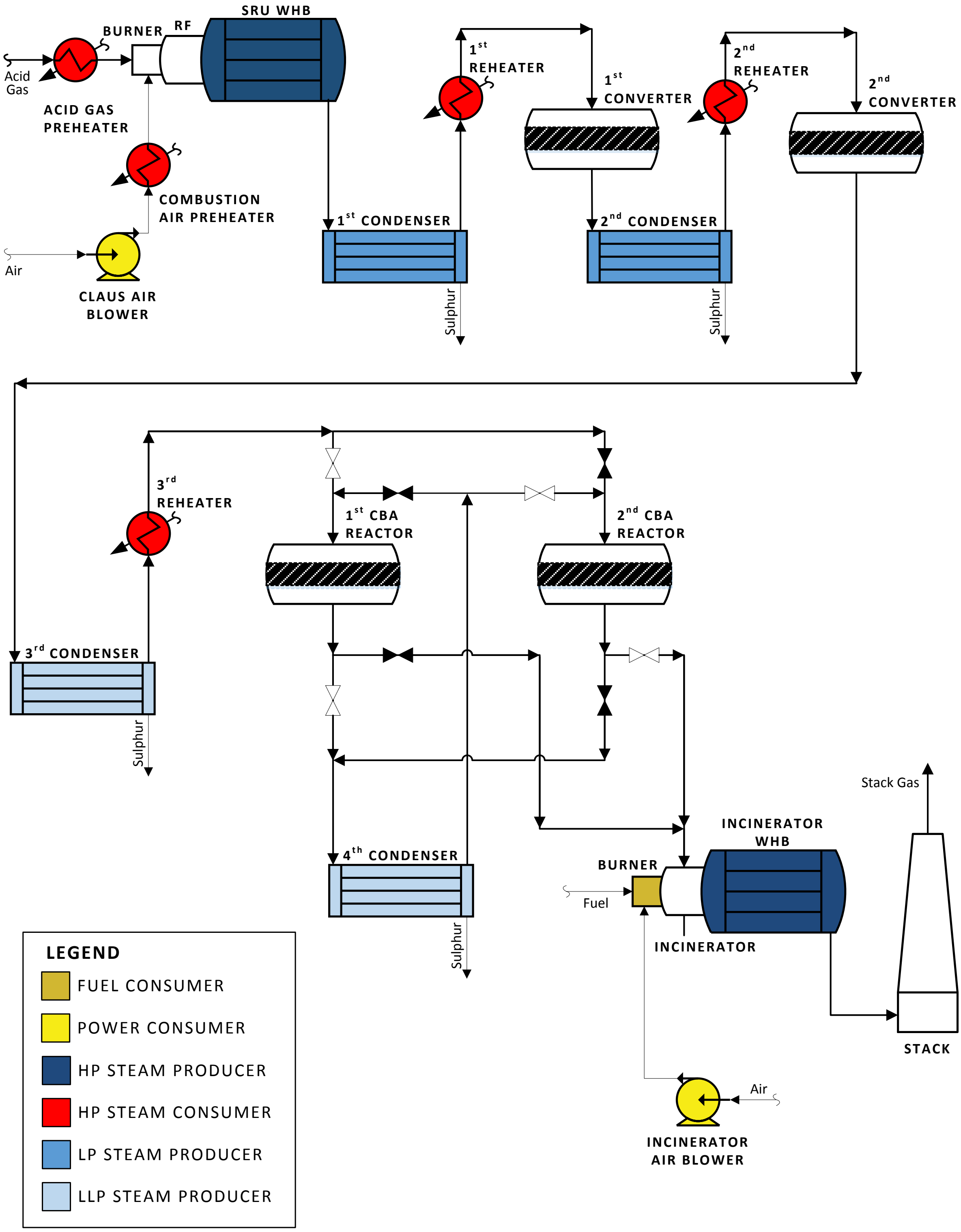


Figure 5. PFD for Benchmark Plant Cases C and D

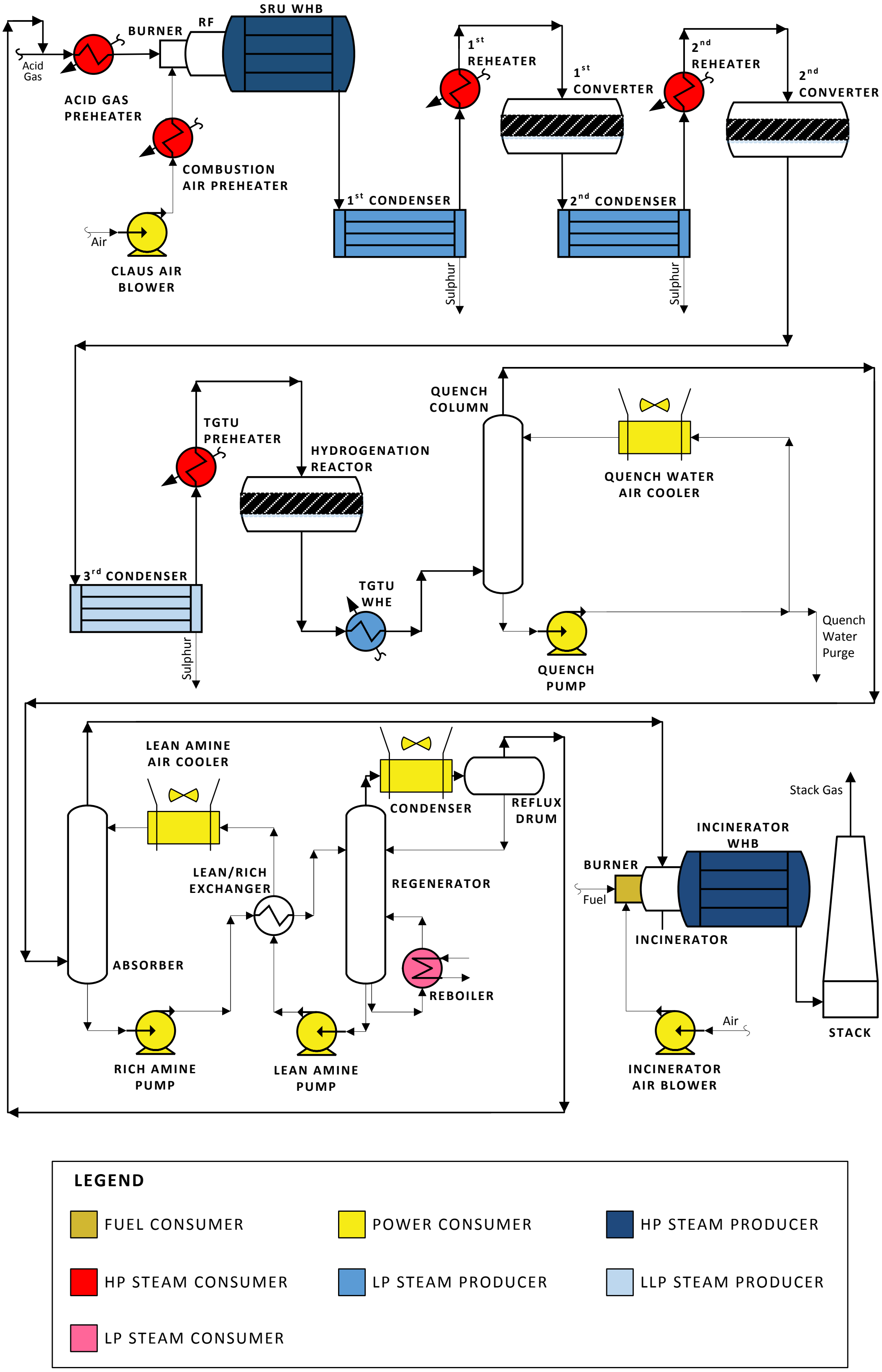
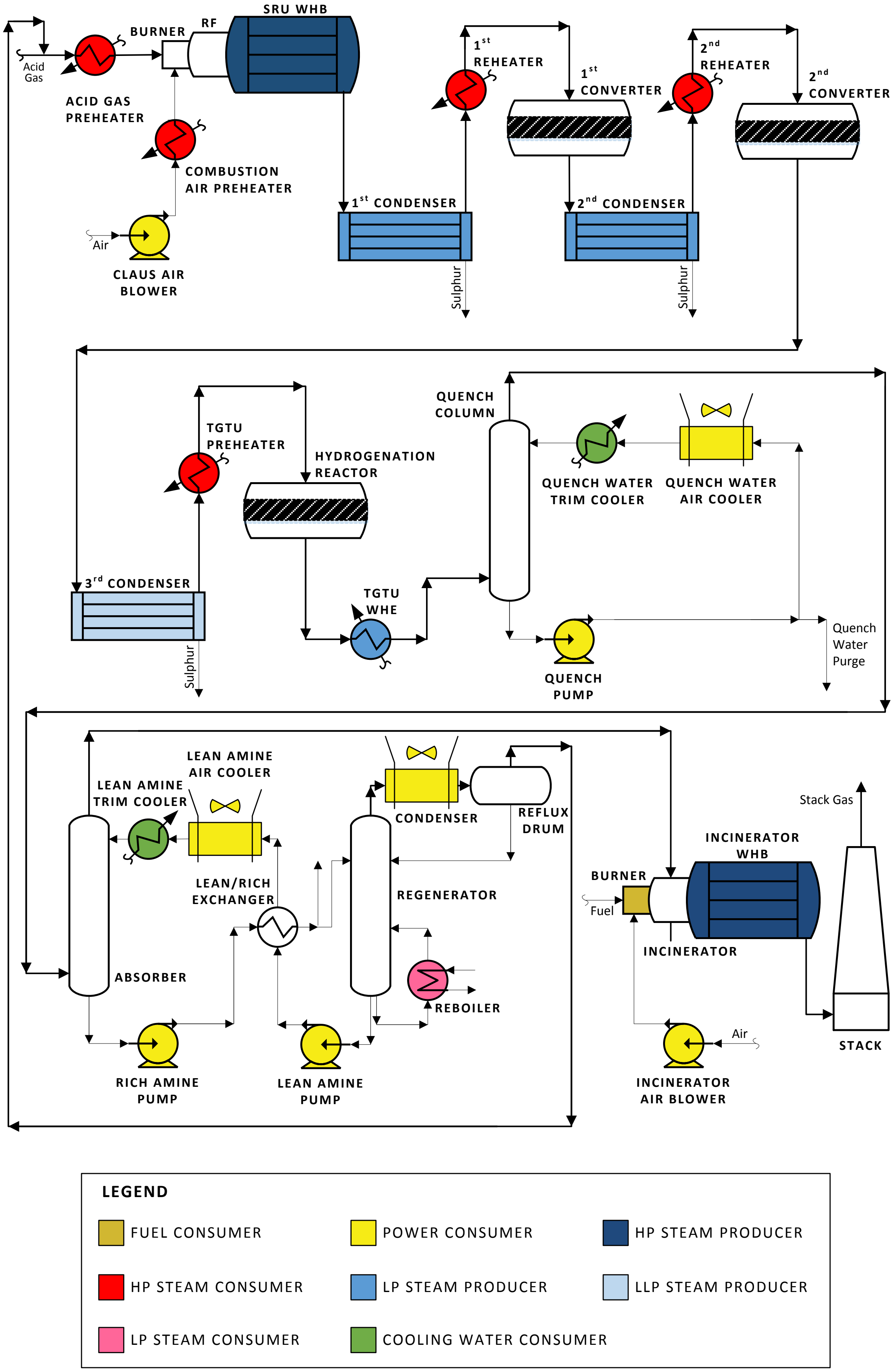


Figure 6. PFD for Benchmark Plant Cases E and F



A standard set of design parameters was employed for all cases to allow relative comparison on a consistent basis. Key design features are aimed at optimizing energy efficiency, as follows:

- Sulphur Recovery Unit
  - Air-only operation, without fuel co-firing
  - Motor-driven Claus air blowers
  - HP saturated steam (40 barg) produced in SRU waste heat boiler (WHB)
  - 2 Claus beds (3 for 97% SRE case) with promoted activated alumina catalyst
  - 2 additional sub-dewpoint beds for 99.0% SRE case
  - LP steam (3.5 barg) produced in 1st & 2nd sulphur condensers
  - LLP steam (1.0 barg) produced in 3rd & 4th sulphur condensers
  - HP saturated steam (40 barg) consumed in SRU preheaters and reheaters

- Amine-based Tail Gas Treatment Unit
  - HP saturated steam (40 barg) consumed in preheater
  - Low temperature hydrogenation catalyst (232 ^oC inlet temperature)
  - LP steam (3.5 barg) produced in TGTU waste heat exchanger (WHE)
  - Lean solvent temperature of 50 ^oC for all except Cases E and F, which was reduced to 40 ^oC to achieve ultra-high SRE (air cooling to 50 ^oC with CW trim cooling)
  - LP steam (3.5 barg) consumed in regenerator reboiler
  - Solvent circulation rate for Case F assumed as 50% of Case E to approximate proprietary solvent[4]

- Incinerator
  - Operated at 815 ^oC (upper limit, required for achieving <5 mg/Nm^3 TRS)
  - 2% excess O_2 in stack gas
  - Fuel fired (LHV of 8,953 $kcal/Nm^3$)
  - Motor-driven incinerator air blowers
  - HP saturated steam (40 barg) produced in incinerator WHB
  - No sulphur pit ejector routed to incinerator

Key process parameters were compared for the range of SRE cases, as summarized in **Table 2**.

Table 2. 1,000 MTPD Benchmark Plant Process Parameters

CASE SRE	A 97%	B 99.0%	C 99.3%	D 99.9%	E 99.98%	F 99.98%
Stack Gas SO_2 (kmol/hr)	38.95	12.97	9.04	1.24	0.28	0.28
Stack Gas CO_2 (kmol/hr)	870.6	885.7	845.4	850.6	851.2	851.2
Total Stack Gas Flow (kmol/hr)	8,012	8,134	6,176	6,188	6,189	6,189
Amine Circulation (m^3/hr)	---	---	176	264	1,026	513

ENERGY BALANCE FOR BENCHMARK PLANT

Thermal energy production/consumption figures for the top producers and consumers in the Benchmark sulphur recovery facility are summarized in **Table 3**, for each of the cases studied. For electric power consumers, the equivalent thermal energy consumption is calculated based on electricity generation, using a steam turbine with an isentropic efficiency of 45% and a generator efficiency of 95%, for an overall thermal to electric energy conversion efficiency of about 43%. This is a change from the previous paper[6], which did not convert electric power to thermal energy, resulting in power energy figures which were roughly 50% of the numbers included in this paper. The overall impact is minor and does not materially impact the results.

It is observed that as SRE increases, energy export decreases, and the facility reverts from net energy export to import at ultra-high recovery efficiency (Case E), primarily due to LP steam consumption for MDEA solvent regeneration in the TGTU. When a highly-selective proprietary solvent is employed (Case F), LP steam consumption is reduced substantially, making the facility closer to energy-neutral.

Table 3. Energy Balance by Top Utility Producers/Consumers for Benchmark Plant (kW)

CASE SRE	A 97%	B 99.0%	C 99.3%	D 99.9%	E 99.98%	F 99.98%
HP STEAM PRODUCERS						
SRU WHB	+53,921	+54,044	+55,256	+55,586	+55,561	+55,561
Incinerator WHB (815°C)	+40,460	+40,615	+30,379	+30,424	+30,429	+30,429
HP STEAM CONSUMERS						
Acid Gas Preheater	-3,994	-3,994	-3,994	-3,994	-3,994	-3,994
Combustion Air Preheater	-3,271	-3,279	-2,900	-2,916	-2,918	-2,918
1st & 2nd SRU Reheaters	-7,376	-7,210	-7,604	-7,638	-7,678	-7,678
3rd SRU Reheater	-2,352	-1,350	---	---	---	---
TGTU Reactor Preheater	---	---	-4,382	-4,401	-4,423	-4,423
LP/LLP STEAM PRODUCERS						
1st & 2nd Sulphur Condensers	+19,544	+19,620	+20,255	+20,353	+20,431	+20,431
3rd Sulphur Condenser	+4,954	+4,092	+5,282	+5,307	+5,329	+5,329
4th Sulphur Condenser	+2,677	+4,185	---	---	---	---
TGTU Hydrogenation WHE	---	---	+3,592	+3,609	+3,622	+3,622
LP STEAM CONSUMERS						
Regenerator Reboiler	---	---	-12,266	-18,610	-86,820	-43,410
MP/LP FUEL CONSUMERS						
Incinerator Burner	-43,838	-47,147	-38,261	-39,426	-39,572	-39,572
MAJOR ELECTRIC POWER CONSUMERS						
Claus Air Blowers	-3,546	-3,556	-4,681	-4,706	-4,711	-4,711
Quench Pumps	---	---	-175	-175	-175	-175
Amine Pumps	---	---	-194	-295	-1,137	-568
Quench Water Air Cooler	---	---	-239	-358	-1,387	-692
Lean Amine Air Cooler	---	---	-159	-246	-641	-320
Regenerator Ovh'd Condenser	---	---	-145	-236	-1,375	-688
Incinerator Air Blowers	-575	-578	-458	-458	-458	-458
MAJOR COOLING WATER CONSUMERS						
Quench Water Trim Cooler	---	---	---	---	-17,366	-8,683
Lean Amine Trim Cooler	---	---	---	---	-9,937	-4,969
ENERGY BALANCE						
Net Energy Import/Export	+56,603	+55,443	+39,306	+31,819	-67,191	-7,889
Comparison to Case A (Net Δ)	---	-2%	-31%	-44%	-219%	-114%

The energy balances for the various cases are summarized by unit operation in **Table 4** and **Figure 7**. These balances clearly illustrate that the SRU is always a net energy exporter whose energy production remains fairly constant, even for Case B, which employs a non-amine-based tail gas treating technology to achieve higher SRE. It is the amine-based TGTU that is responsible for increasing energy consumption as SRE increases.

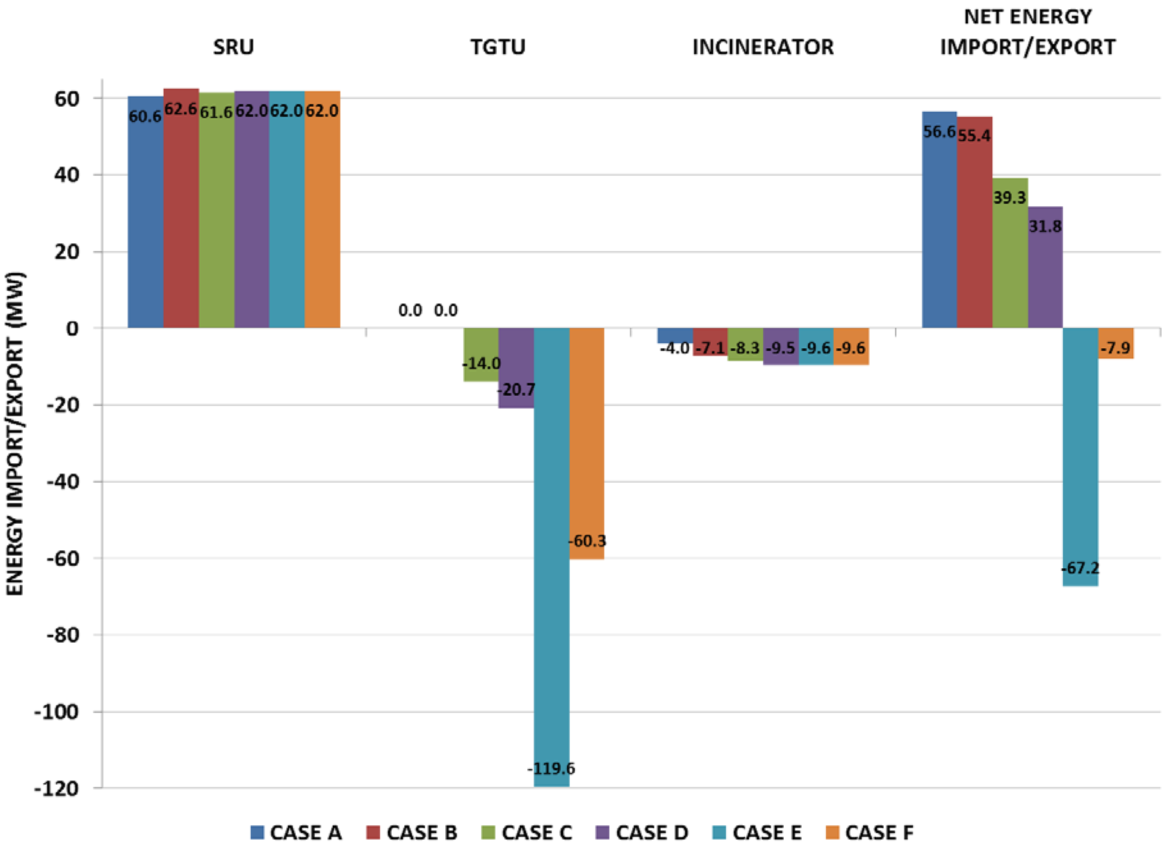
Highly-selective solvents can offer reductions in LP steam and solvent cooling requirements but the overall impact of the TGTU on sulphur recovery energy production is still significant. The world average sulphur recovery efficiency for new plants is around 99.9% (Case D), which results in a 44% energy penalty on the standalone Claus plant, as shown in **Tables 3** and **4**. Despite its detrimental impact on the overall SRU/TGTU energy balance, the amine-based tail gas treating process is currently the only conventional technology available for achieving guaranteed SRE in excess of about 99.3%.

Table 4 also shows that incinerator energy consumption increases slightly as SRE increases, simply due to a reduced quantity of H_2S in the tail gas. This results in lower tail gas heating value and therefore greater fuel requirements, although the overall impact is marginal.

Table 4. Benchmark Plant Energy Balance by Processing Unit (kW)

CASE SRE	A 97%	B 99.0%	C 99.3%	D 99.9%	E 99.98%	F 99.98%
SRU	+60,557	+62,552	+61,614	+61,992	+62,020	+62,020
TGTU	---	---	-13,968	-20,712	-119,610	-60,307
INCINERATOR	-3,953	-7.110	-8,340	-9,460	-9,601	-9,601
NET	+56,603	+55,443	+39,306	+31,819	-67,191	-7,889

Figure 7. Benchmark Plant Energy Balance by Processing Unit



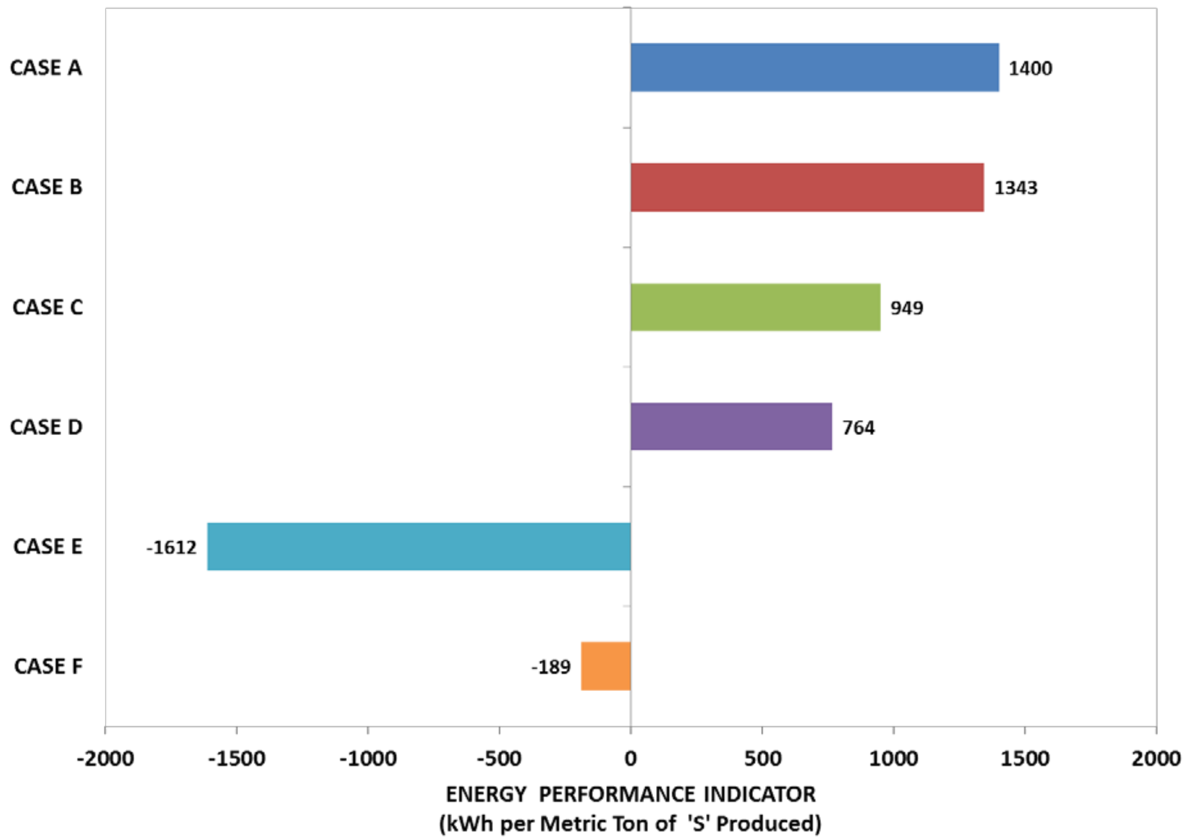
ENERGY KEY PERFORMANCE INDICATORS (KPIs) FOR BENCHMARK PLANT

The net energy balance figures provided in **Tables 3** and **4** are converted to "thumb rule" targets that can be used to assess a sulphur recovery facility's energy performance, as provided in **Table 5** and **Figure 8**. These key performance indicators (KPIs) can be utilized by operators to evaluate whether their facilities are operating in accordance with best energy efficiency standards. While such metrics may not have been viewed as particularly important previously, the authors are observing an increasing trend of operators wishing to make the best use of the sulphur recovery unit's energy benefits. As a result, SRU/TGTU energy efficiency is being increasingly evaluated and scrutinized, particularly for those sour facilities with relatively large sulphur recovery requirements.

Table 5. Energy Performance KPIs for Benchmark Plant

CASE	A	B	C	D	E	F
SRE	97%	99.0%	99.3%	99.9%	99.98%	99.98%
kWh per Metric Ton 'S' Produced	+1,400	+1,434	+949	+764	-1,612	-189
kWh per Nm^3 H_2S in Acid Gas Feed	+1.94	+1.90	+1.35	+1.09	-2.31	-0.27

Figure 8. Energy Performance KPIs (kWh per Metric Ton of 'S' Produced)



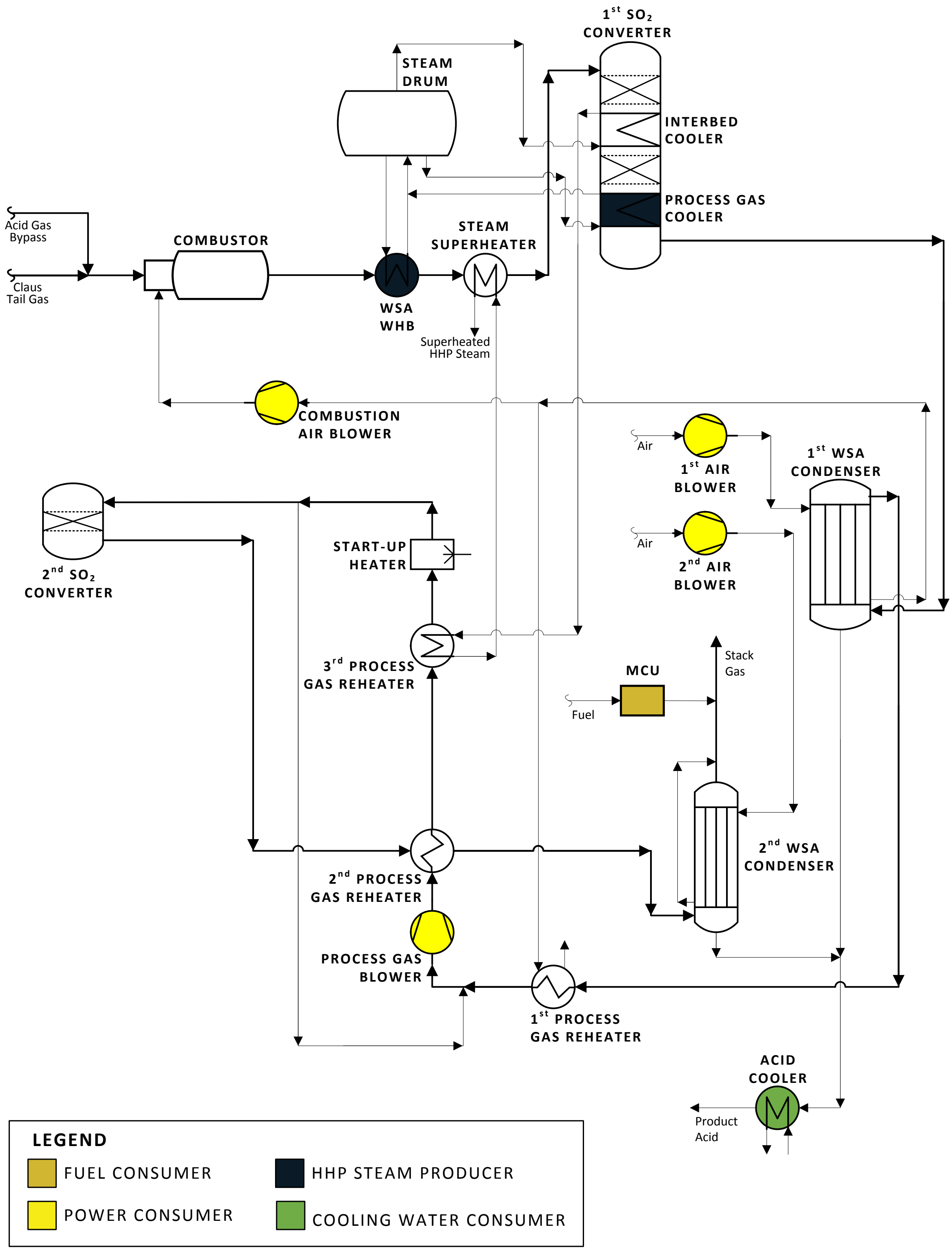
It is important to keep in mind specific feed conditions and plant design configuration when applying this information. A different configuration, feedstock and/or operating philosophy can lead to significant variations in KPIs, as we will see in the Case Study which follows. For example, a plant which is equipped with an incinerator WHB can generate up to 40% more HP steam than one that is not, as illustrated in **Table 3**. Some other examples that can lead to widely varying KPIs include fuel gas co-firing in the SRU, the use of an RGG in the TGTU, installation of low-temperature catalyst in the TGTU and TGTU solvent chilling requirements, to name a few.

For the most part, the energy efficiency of the Benchmark Plant design has been optimized across all Cases A-F, with the exception of incinerator operating temperature, which could be reduced by around 165 ^oC, depending on the emission regulations. However, since an incinerator WHB is employed in the Benchmark Plant design, additional waste heat is recovered at the higher temperature and the overall impact on energy efficiency is negligible. There are some other minor opportunities for improving energy efficiency that are not included in the Benchmark Plant design, such as BFW preheat in the final condenser and sulphur cooler to maximize HP steam production (rather than generating LLP steam), but the overall impact on energy export/import for these items is not expected to significantly impact the results of this study.

TAIL GAS TREATMENT WITH WET SULPHURIC ACID (WSA) TECHNOLOGY

As illustrated in **Figure 7,** the amine-based TGTU has a negative impact on the overall energy balance of the facility, which becomes quite significant as SRE increases. Thus, an alternate tail gas treating technology has been considered for comparison. In the Wet Gas Sulphuric Acid (WSA) process, residual sulphur in the Claus tail gas is recovered as concentrated sulphuric acid. The oxidation process is exothermic and no solvent regeneration is required; hence, energy performance should be improved, versus the amine-based tail gas treating process. A potential downside is the production of two separate products; however, this may be outweighed by the energy benefits in certain scenarios. A process flow diagram for a typical WSA unit, indicating the top utility producers/consumers, is provided in **Figure 9**.

Figure 9. PFD for Wet Gas Sulphuric Acid (WSA) Tail Gas Treating Cases G and H



The WSA process consists of 7 main steps:

1. Combustion of tail gas

   Claus tail gas, along with a portion of the acid gas which is bypassed around the SRU, are combusted with preheated air from the combustion air blower. Acid gas bypass is required to ensure temperatures sufficient for complete combustion of tail gas sulphur species and to avoid continuous fuel gas consumption in this burner.

2. Gas cooling and steam generation

   Heat is recovered downstream of the combustion process as saturated and superheated HHP steam at 58.5 barg, as the process gas is cooled to approximately 425 ^{o}C.

3. First step of SO_2 oxidation and subsequent cooling

   Process gas from the waste heat boiler is introduced to the 1st SO_2 converter, which is equipped with two catalytic conversion stages. SO_2 is oxidized to SO_3 according to the exothermic reaction in Equation 4.

   $$SO_2 + 1/2O_2 \rightarrow SO_3 \qquad \Delta H = -24 \text{ kcal/mol} \qquad (4)$$

   The reaction is a temperature and concentration dependent equilibrium reaction. Therefore, to optimize the temperature approach to equilibrium and achieve a high rate of conversion, the process gas is cooled in between the catalyst beds in the interbed cooler. Downstream of the second catalyst bed, the process gas is cooled once more in the process gas cooler to 290 °C before it enters the 1st WSA condenser.

4. Acid condensation in 1st WSA condenser

   In the 1st WSA condenser, process gas is further cooled with ambient air supplied by the 1st air blower and part of the SO_3 is hydrated to H_2SO_4 vapour according to the exothermic reaction in Equation 5.

   $$SO_3 + H_2O \rightarrow H_2SO_4(g) \qquad \Delta H = -25 \text{ kcal/mol} \qquad (5)$$

   Acid is condensed inside the vertical glass tubes and collected at approximately 260 °C at the bottom of the condenser. Cooling air leaves the 1st WSA condenser at a temperature above 240 °C and is used as combustion air in the WSA combustor, for reheating the process gas from the 1st WSA condenser, for mixing with the cooling air at the inlet of the WSA condensers, and for BFW preheating.

5. Process gas heating

   Process gas leaving the 1st WSA condenser is preheated to approximately 405 ^{o}C in four to five steps (see **Figure 9**) before entering the 2nd SO_2 converter.

6. Second step of SO_2 oxidation and acid condensation

   After preheating, process gas enters the 2nd SO_2 converter, comprised of one catalytic stage, where any remaining SO_2 is oxidized to SO_3. Process gas leaving the 2nd SO_2 converter is then cooled in the 2nd process gas heater before entering the 2nd WSA condenser.

   In the 2nd WSA condenser, process gas is cooled further with ambient air supplied by the 2nd combustion air blower and all of the SO_3 is hydrated to H_2SO_4. Acid is condensed inside vertical glass tubes and collected at approximately 190 °C at the bottom of the condenser. Process gas leaving the 2nd WSA condenser is routed to the stack. Cooling air leaves the condenser at a temperature above 180 °C and is used for conditioning of the stack gas.

7. Acid concentration and cooling

   If no special precautions are taken, part of the H_2SO_4 vapour will condense as fine mist particles that are too small to be separated from the process gas in the WSA condensers. To minimize the formation of this mist in the outlet gas, a mist control unit (MCU) is installed. In the MCU, silicone oil is combusted on a continuous basis with fuel gas to generate a small hot flue gas stream containing very small silicone oxide particles. These particles act as nuclei on which the acid mist can agglomerate to form larger droplets until they are large enough to enable proper separation from the process gas in the WSA condensers.

   Concentrated sulphuric acid (96-97 wt%) collected at the bottom of the WSA condensers is cooled to 70 °C by recirculation before final cooling to about 40 °C in a water-cooled plate type heat exchanger. Product acid offtake originates from a side stream at the outlet of the acid cooler.

ENERGY PERFORMANCE COMPARISON OF WSA PLANT

Two cases for Claus tail gas treatment via the WSA process have been considered for energy performance comparison, using the same acid gas feed flow and composition as the Benchmark Plant, and an SRE of 99.9%. Case G features a 1-bed Claus plant followed by a WSA unit, while Case H features a 2-bed Claus plant followed by a WSA unit. Both cases require acid gas bypass around the SRU to avoid continuous fuel gas consumption in the tail gas burner and to avoid the production of dilute acid. In Case H, the bypass requirement is slightly higher due to a lower concentration of sulphur species in the Claus tail gas when 2 Claus beds are employed. Because the WSA process produces higher pressure steam than the SRU (55 vs. 40 barg), this utility service is referred to as high high pressure (HHP) steam (containing 128°C of superheat) and thermal energy content is calculated accordingly.

Key process parameters for the Claus + WSA tail gas treating cases are summarized in **Table 6.**

Table 6. 1,000 MTPD Benchmark Plant Design/Process Parameters for WSA Cases

CASE SRE	G 99.9%	H 99.9%
Number of Claus Stages	1	2
Acid Gas Bypass (%)	35	40
Total Stack Gas Flow (kmol/hr)	9,637	9,235
SO_x Emission, as SO_2, 3% O_2, Wet (mg/Nm^3)	430	430
Elemental Sulphur Production (TPD)	537	574
% of Feed H_2S Converted to Elemental Sulphur	54%	57%
Sulphuric Acid Production, 96-97 wt% (TPD)	1,468	1,352
% of Feed H_2S Converted to Sulphuric Acid	46%	43%
Superheated HHP Steam Export, 55 barg, 400°C (t/h)	127	122

It is important to note that in both WSA cases, the sulphur recovery facility will convert slightly less than half of the H_2S in the acid gas feed to sulphuric acid, rather than elemental sulphur. This will have to be accommodated when considering product storage, transportation and marketing requirements, which will obviously have an impact on capital and operating costs; however, the energy impact will be minimal and will not affect the outcome of this evaluation. Depending on the location of the facility, marketing limitations may preclude a sulphuric-acid-producing option from being considered.

ENERGY BALANCE FOR WSA PLANT

Thermal energy production/consumption figures for the top producers and consumers in the WSA Plant are summarized in **Table 7**, for each of the cases studied. Equivalent thermal energy consumption by electric power consumers has been calculated based on electricity generation using a steam turbine, with conversion efficiencies similar to the previous cases examined.

Compared to Case D of the Benchmark Plant, which uses an amine-based TGTU to achieve a similar SRE, the WSA cases can reach the same extent of sulphur recovery with more than 3 times the net energy export. The primary reasons for this are the avoidance of continuous fuel consumption in the incinerator, combined with improved efficiency in the tail gas combustion heat recovery system, as well as elimination of the requirement for LP steam for solvent regeneration. The WSA cases consume more electric power due to the additional air and process gas blowers, and also have some additional energy requirements for acid cooling; however, the net result is that the WSA cases improve the energy efficiency of the facility.

Table 7. Energy Balance by Top Utility Producers/Consumers for 99.9% SRE Cases (kW)

CASE SRE	D 99.9%	G 99.9%	H 99.9%
SUPERHEATED HHP STEAM PRODUCERS			
WSA WHB & Process Gas Cooler	---	+97,409	+93,574
HP STEAM PRODUCERS			
SRU WHB	+55,586	+36,131	+33,352
Incinerator WHB (815°C)	+30,424	---	---
HP STEAM CONSUMERS			
Acid Gas Preheater	-3,994	-2,596	-2,396
Combustion Air Preheater	-2,916	-1,895	-1,750
1st SRU Reheater	-6,061	-3,940	-3,637
2nd SRU Reheater	-1,577	---	-946
TGTU Reactor Preheater	-4,401	---	---
LP/LLP STEAM PRODUCERS			
1st & 2nd Sulphur Condensers	+20,353	+13,229	+12,212
3rd Sulphur Condenser	+5,307	---	+3,184
TGTU Hydrogenation WHE	+3,609	---	---
LP STEAM CONSUMERS			
Regenerator Reboiler	-18,610	---	---
MP/LP FUEL CONSUMERS			
Incinerator Burner	-39,426	---	---
WSA MCU	---	-42	-42
MAJOR ELECTRIC POWER CONSUMERS			
Claus Air Blowers	-4,706	-3,059	-2,824
Quench Pumps	-175	---	---
Amine Pumps	-295	---	---
Quench Water Air Cooler	-358	---	---
Lean Amine Air Cooler	-246	---	---
Regenerator Ovh'd Condenser	-236	---	---
Incinerator Air Blowers	-458	---	---
WSA Blowers	---	-17,143	-16,508
MAJOR COOLING WATER CONSUMERS			
WSA Acid Cooler	---	-7,318	-6,745
ENERGY BALANCE			
Net Energy Import/Export	+31,819	+110,777	+107,474
Comparison to Case D (Net Δ)	---	+248%	+238%

The energy balance for the WSA Plant cases is summarized by unit operation in **Table 8** and by utility system in **Table 9**. Net energy export from the SRU is reduced in the WSA cases due to acid gas bypass; however, the tail gas treating section of the process is converted from an energy consumer to an energy producer when WSA is employed. There is a marginal increase in overall electric power consumption in the WSA cases but the net result is that energy export from the WSA cases is more than three times greater than the amine-based TGTU case.

Table 8. Energy Balance by Processing Unit for 99.9% SRE Cases (kW)

CASE SRE	D 99.9%	G 99.9%	H 99.9%
SRU	+61,992	+37,870	+37,195
TGTU	-20,712	0	0
INCINERATOR	-9,460	0	0
WSA	0	+72,907	+70,279
NET	+31,819	+110,777	+107,474

Table 9. Energy Balance by Utility System for 99.9% SRE Cases (kW)

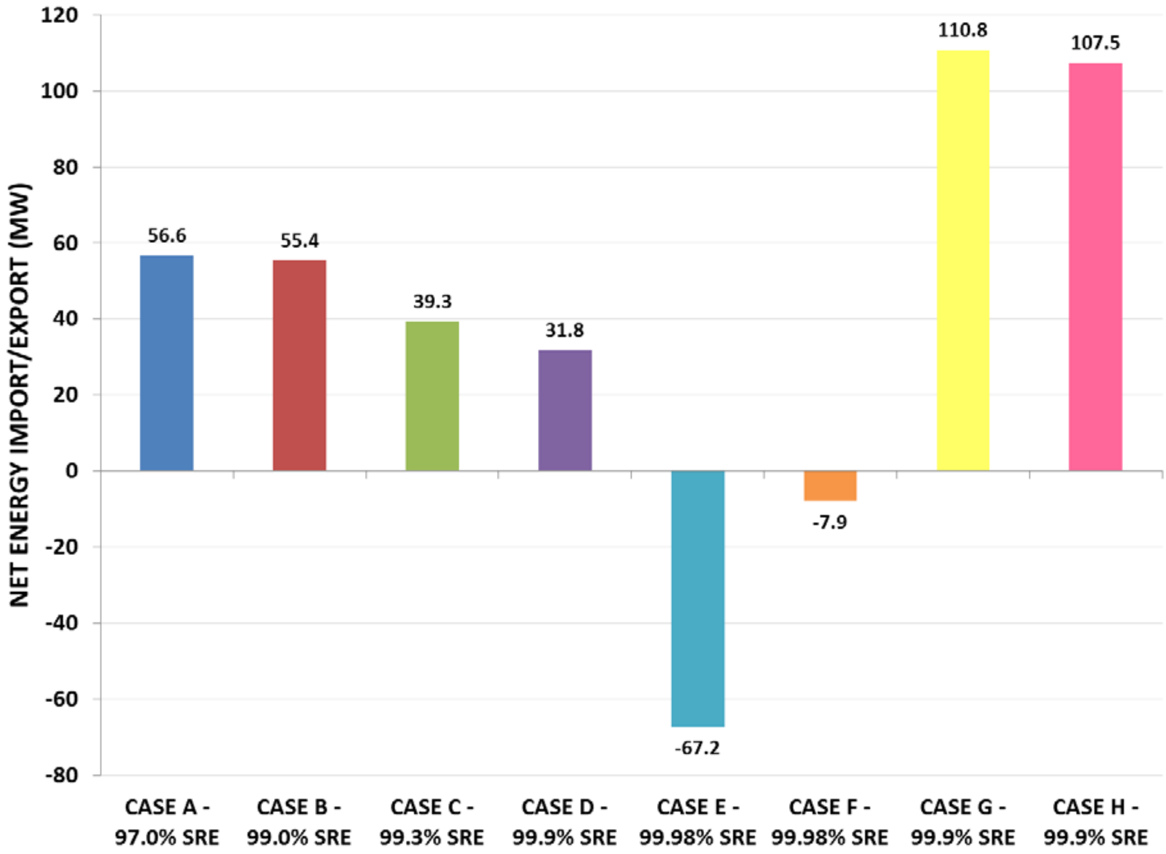
CASE SRE	D 99.9%	G 99.9%	H 99.9%
HHP STEAM	0	+97,409	+93,574
HP STEAM	+67,061	+27,699	+24,623
LP STEAM	+10,659	+13,229	+15,396
FUEL GAS	-39,426	-42	-42
ELECTRIC POWER	-6,475	-20,202	-19,332
COOLING WATER	0	-7,318	-6,745
NET	+31,819	+110,777	+107,474

A comparison of the net energy import/export across all of the sulphur recovery technologies and SREs considered is provided in **Figure 10**. In comparing Case C to Cases A and B, it is apparent that the energy-consuming, amine-based tail gas treating unit erodes the energy benefits of the SRU. As SRE increases (Cases D and E), energy requirements for the TGTU also increase, resulting in an even greater negative impact on the overall energy balance. The use of a proprietary solvent in the TGTU (Case F) reduces TGTU energy consumption but it is still a net energy consumer. Because of the fact that the WSA tail gas treating facility is a significant energy producer, the energy balance of the overall sulphur recovery facility is enhanced to nearly double the SRU energy export (Cases G and H).

It should be noted that additional product storage, transportation and marketing costs are likely to be incurred in the WSA cases, which produce two products (elemental sulphur and H_2SO_4) rather than one. While this will result in additional capital and operating costs, preliminary calculations indicate that the overall net present cost for the WSA cases will still be more attractive, assuming that there is a local market for the product acid.

Currently, WSA tail gas treating technology has been compared to an amine-based TGTU achieving 99.9% SRE (Case D). It would be possible to achieve 99.98% SRE (Cases D and E) utilizing WSA technology, with only a marginal impact on energy consumption. Due to time constraints, this scenario was not qualitatively assessed; however, the energy export figures are expected to be only slightly lower than Cases G and H.

Figure 10. Net Energy Import/Export for All Cases



ENERGY KEY PERFORMANCE INDICATORS (KPIs) FOR WSA PLANT

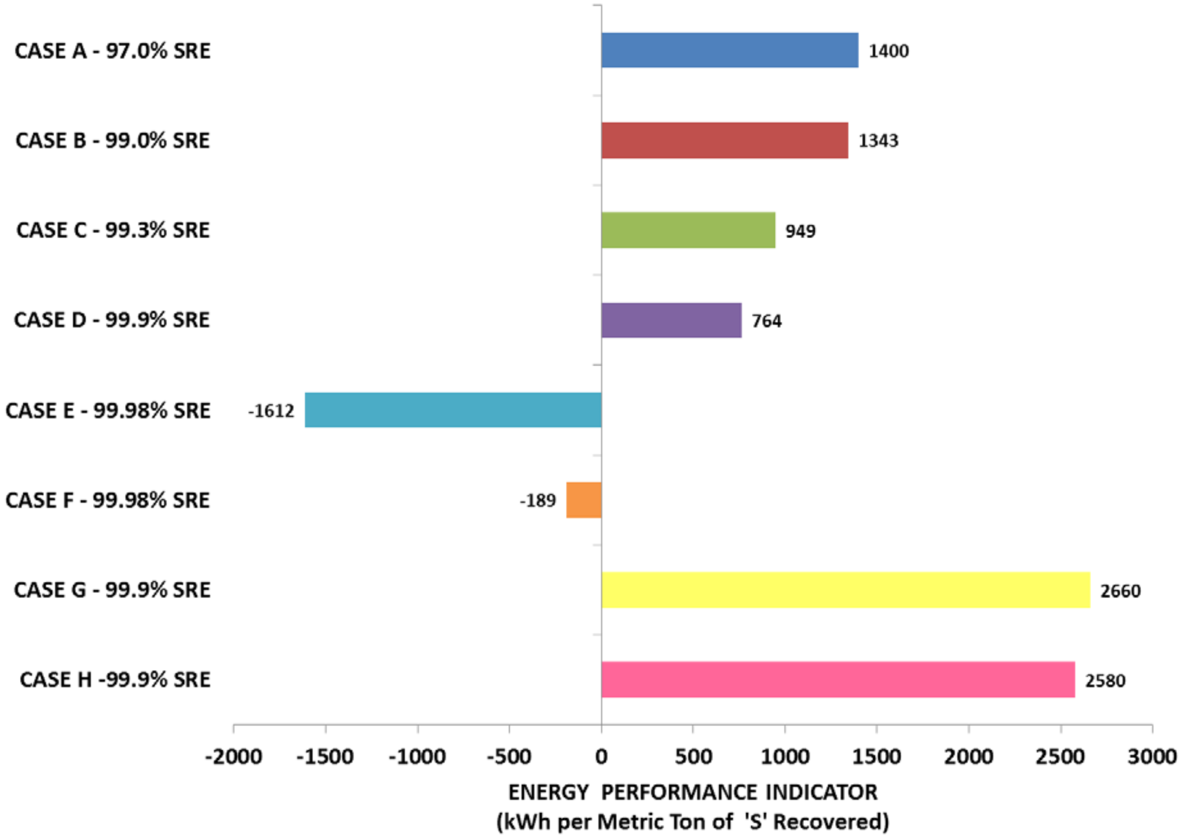
The net energy balance figures provided in **Table 7** are converted to energy performance KPIs, as provided in **Table 10**. For the WSA cases, these KPIs are determined per metric ton of sulphur recovered, whether as elemental sulphur or sulphuric acid.

Table 10. Energy Performance KPIs for 99.9% SRE Cases

CASE	D	G	H
SRE	99.9%	99.9%	99.9%
kWh per Metric Ton 'S' Recovered	+764	+2,660	+2,580
kWh per Nm^3 H_2S in Acid Gas Feed	+1.09	+3.80	+3.69

A comparison of the energy performance KPIs across all of the sulphur recovery technologies and SREs considered is provided in **Figure 11**. The results are proportional to those provided and discussed in **Figure 10**.

Figure 11. Energy Performance KPIs for All Cases (kWh per Metric Ton of 'S' Recovered)



ENERGY EXPORT POTENTIAL FROM UAE SULPHUR RECOVERY FACILITIES

It can be interesting to apply the benchmark plant energy balance data to real-world scenarios, in an attempt to estimate the actual and potential energy benefits that can be realized from sulphur recovery facilities. In 2016, the Middle East became the world's largest sulphur producing region. Current UAE sulphur production is approximately 19,000 MTPD (7 MMTPA), equivalent to roughly 11% of global production. As indicated in **Table 5**, for every ton of sulphur produced, a Claus SRU is capable of exporting approximately 1,400 kWh of thermal energy. Considering a thermal energy to electric power conversion efficiency of 43%, current mechanical power generation potential from all UAE SRUs (without tail gas treating) is in the range of 500 MW.

As indicated by the Benchmark Plant analysis in **Table 3**, amine-based tail gas treating may consume between 40 and 100% of the energy produced by the SRU. For a 99.9% SRE facility utilizing a proprietary solvent, the TGTU should consume roughly 50% of the energy produced by the SRU. Thus, **the overall power generation from UAE SRUs should currently be in the range of 250 MW** (or greater, as not all UAE SRUs are equipped with amine-based tail gas treaters). This energy would be extremely beneficial for supplementing the power requirements of the sour gas processing facilities in which they are installed. However, it is important to note that certain factors may erode these SRU energy benefits in some cases, especially in facilities processing lean acid gas in hot climates, as is the case with most sour gas plants in the Middle East. Factors which may diminish SRU energy benefits include fuel gas co-firing, refrigeration requirements in the TGTU, employment of an RGG rather than steam-heated TGTU preheater, and absence of an incinerator WHB, to name a few.

Within approximately 10-15 years, UAE sulphur production could double, which presents the possibility for producing an additional 250 MW (500 MW total) of power from all UAE SRUs, and potentially more, if technologies which maximize energy efficiency are employed. Currently there is only a very limited local market for sulphuric acid production; however, if regional development leads to eventual local acid demand, future tail gas treating facilities might consider employing WSA technology as a means of reducing the overall energy import requirements of gas treating facilities. Based on the analysis provided in this paper, if 50% of future tail gas treating facilities were designed for sulphuric acid production (resulting in up to 15,000 MTPD H_2SO_4), the total power export from all sulphur recovery facilities could increase from 500 MW to approximately 750 MW.

CONCLUSIONS

Sulphur recovery facilities provide significant energy benefits and should be leveraged to their fullest potential via astute design and optimized operation, deliberately focused on energy conservation. This is especially important in the current climate of low oil price and reduced margins. Conventional tail gas treating technologies can significantly erode energy benefits provided by Claus SRUs and therefore should be designed and operated to achieve lowest acceptable SRE (highest acceptable stack gas SO_2 content). Given the huge energy requirements of amine-based TGTUs, there may be a case to lobby for relaxed SO_2 emissions regulations for future facilities.[5] However, this is a lofty goal, and should it prove unattainable, alternate technologies may be considered. In the case of WSA, the production of an additional sulphuric acid product would require careful consideration in terms of storage, transportation and marketing.

NOMENCLATURE

°C	degrees Celsius
°F	degrees Fahrenheit
AGE	acid gas enrichment
AGRU	acid gas removal unit
barg	bar gauge
BFW	boiler feed water
BTEX	benzene, toluene, ethylbenzene, xylene
BTU	British thermal unit
C_1	methane
CBA	Cold Bed Adsorption
CO	carbon monoxide
CO_2	carbon dioxide
CW	cooling water
h	hour
H_2O	water
H_2S	hydrogen sulphide
H_2SO_4	sulphuric acid
HHP	high pressure
HP	high pressure
kcal	kilocalorie
kmol	kilomole
KPI	key performance indicator
kW	kilowatt
kWh	kilowatt-hour
LHV	lower heating value
LLP	low pressure
LP	low pressure
m^3	cubic metres
MCU	mist control unit
MDEA	methyl diethanolamine
mg	milligram
MMTPA	million metric tons per annum
mol	mole
mol%	mole percent
MP	medium pressure
MTPD	metric tons per day
MW	megawatt
Nm^3	normal cubic metres
O_2	oxygen
PFD	process flow diagram
RF	reaction furnace
RGG	reducing gas generator
S	elemental sulphur
SO_2	sulphur dioxide
SO_3	sulphur trioxide
SRE	sulphur recovery efficiency
SRU	sulphur recovery unit
S_x	elemental sulphur
t	ton
TGTU	tail gas treating unit
TRS	total reduced sulphur
UAE	United Arab Emirates
vol%	volume percent
WHB	waste heat boiler
WHE	waste heat exchanger
WSA	Wet Gas Sulphuric Acid
wt%	weight percent
Δ	change
ΔH	enthalpy of reaction

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