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Optimising heat recovery with CPHEs

Compact plate heat exchangers are an efficient way to optimise heat recovery in refineries, especially when installation space or weight is crucial and exotic materials are needed to deal with corrosion issues

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In today's refinery business, there is an increasing emphasis on expanding capacity while reducing energy consumption and maintaining a positive environmental profile. The inherent properties of compact plate heat exchangers (CPHEs) make them a more efficient and compact solution to traditional shell-and-tube heat exchangers (S&Ts). CPHEs make it possible to achieve increased capacity and to recover more heat using fewer heat exchangers, which are also much smaller and can easily be fitted within the footprint of existing installations. The greater energy recovery results in savings in fuel consumption as well as reductions in emissions, giving investment payback times that are typically very attractive.

Satisfied operators also confirm that the use of CPHEs provides faster response to changes in the process, such as at plant startup and shutdown, in addition to longer intervals in between services. When service is needed, unrestricted access to the heat-transfer surfaces means it is easier to restore full heat-transfer efficiency.

Heat-transfer efficiency

The main feature differentiating the CPHE from the traditional S&T is the use of corrugated plates to form the heat exchanger channels. When these plates are stacked, the many contact points force the fluid to spiral its way through the channels, thereby inducing high turbulence (Figure 1). This means that for the same flow velocity through the channels, a CPHE achieves greater turbulence than a S&T, thus giving rise to thermal efficiency that is three to five times higher.

This greater turbulence leads to a higher wall shear stress, permitting the CPHE to operate for longer intervals with no need for maintenance. This is because the wall shear stress has a cleaning effect that reduces any fouling inside the heat exchanger.

Heat-recovery possibilities

Another important feature of the CPHE is that it can operate with a counter-current flow (ie, the hot fluid enters the heat exchanger at the end where the cold fluid exits). This makes it possible to handle crossing-temperature programs (where the cold fluid is heated to a temperature that is higher than the outlet temperature of the hot fluid) in a single heat exchanger (Figure 2). This is especially important in heat recovery, where the cold fluid can be heated to temperatures very close to those of the hot fluid, hence recovering as much energy as possible.

The temperature difference between the hot and the cold fluids (mean temperature difference — MTD) acts as the driving force for heat transfer. The larger the MTD, the more effortless the heat transfer, and vice versa. The effort needed to carry out a certain heat-transfer duty is often measured in terms of its NTU value (theta [θ] value or thermal length). For single phase service, this parameter is calculated as:

Equation 1: $$\text{NTU} = \theta = \frac{T_1 - T_2}{\text{MTD}}$$

T_1 = inlet temperature
T_2 = outlet temperature

For heat-recovery duties, the temperature program (the difference between the inlet temperature and outlet temperature) is normally large and/or the temperature difference between the two fluids (MTD) is very small. Equation 1 shows that this results in a large NTU value. In reality, this means the driving force for the heat-recovery duty is low and the two fluids have to remain in contact for a long time for them to exchange heat.

In a S&T, this is tackled by making the tubes longer, arranging the tubes with many passes and/or connecting several tubes in series. This often results in hydraulic problems, because the channel velocity through the large units is reduced, thereby lowering the thermal efficiency of the heat exchanger even more (as well as increasing the fouling problems).

In contrast, the high thermal efficiency of the CPHE, combined with opportunities for operating with a counter-current flow, allows the CPHE to deal with long temperature programs with a small MTD. As a result, in many

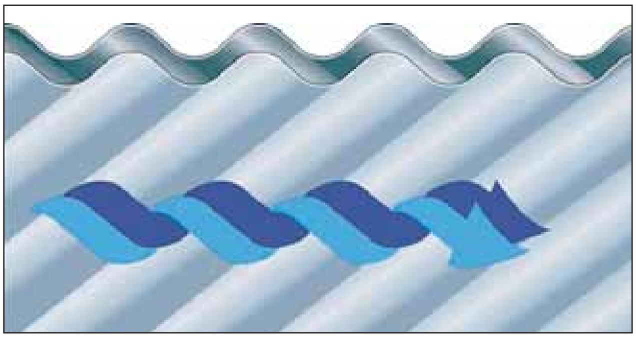

***Figure 1** Turbulent flow through heat exchanger channels, generated by corrugated plates*

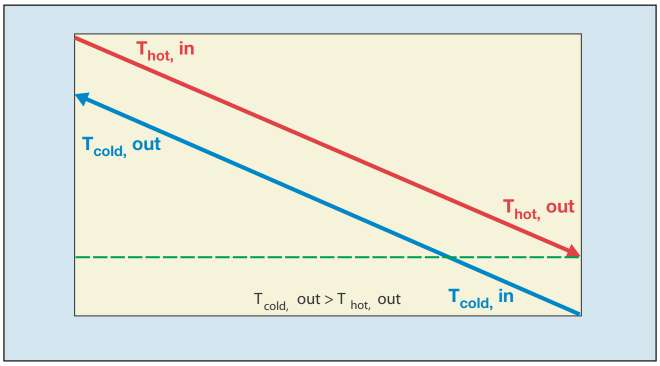

***Figure 2** Crossing-temperature program*

Comparison of CPHE and S&T design results in duties of various degrees of heat recovery

	Diesel/crude HX		Diesel stripper feed/bottom HX	
	Diesel 173 ->88°C NTU = 2.0 Crude 89 <- 70°C NTU = 0.45		Bottoms 251 -> 83°C NTU = 6.0 Feed 215 <- 62°C NTU = 5.5	
	CPHE	S&T	CPHE	S&T
Number of heat exchangers	1	1	3 (parallel)	4 (series)
Number of passes	11/4	4/1 (tube/shell)	16/16	4/1 (tube/shell)
Pressure drop, kPa	85/115	60/82	90/120	90/115
Shear, Pa	31/112	4 (tube)	16/21	3.2 (tube)
Total HTA, m^2	70	400	770	3344

Table 1

Comparison of CPHE and S&T

	CPHE	S&T
Wall shear stress	5–8	1
Heat transfer efficiency	3–5	1
Heat transfer area	1	3–5
Pressure drop	1.2–1.5	1
Temperature pinch	3–5°C	N/A
Footprint	1	5–10
Service area	1	4–10
Weight empty	1	1.5–4
Weight full	1	2–5
Installation cost	1	1.5–2
Hold-up volume	1	30–40

Table 2

cases, only a single CPHE is needed to tackle the required heat-recovery duties, as shown in Table 1.

Installation concerns

As can be seen in Figure 3, the smaller heat-transfer area (HTA) is not the only source of savings. Since the CPHE features such a compact basic design, the HTA required for each specific duty can also be assembled in an effective way. This means a CPHE with around 320m^2 of HTA needs less than 1.5m^2 of floor space for the installation and around 10m^2 of total floor space, including the service area (with 1m added all around the equipment). A corresponding S&T heat exchanger with 6m-long tubes would need around 15m^2 of floor space for the installation and 60m^2 of floor space including the service area, because room for removing the tube bundle must be taken into consideration.

The compactness of the CPHE design also accounts for a reduced weight, which cuts down on installed cost (investment plus installation), especially when construction and/or foundation work is needed. When estimating the installed cost, a factor of 3.0–3.5 times the initial investment cost is often used for S&Ts, while for CPHEs the corresponding factor is normally less than 2.0. In addition, the reduced hold-up volume means the CPHE responds much faster to any changes in the process operating parameters, such as at startup and shutdown. Table 2 summarises the benefits of CPHEs relative to S&Ts.

Service needs

As has already been pointed out, the greater turbulence and the elimination of hydraulic problems extend the operating intervals between services. Service work is also done more quickly when needed, which saves money on both maintenance costs and production downtime. With the better response to process changes, the plant can be shut down and restarted more quickly. If chemical cleaning is used, the lower hold-up volume makes this process faster, with fewer chemicals to dispose of once the cleaning is complete.

If mechanical cleaning is needed, simply unbolting the frame provides complete access to the heat-transfer surface for cleaning using a hydro jet of up to 500 bar, as shown in Figure 4.

Heat recovery in crude preheat trains

In a refinery, the one process that requires the largest amount of energy is the preheating of the crude oil. This is therefore where most of the gains from using CPHEs for heat recovery can be made. However, as will be discussed in further detail, CPHEs can also bring benefits to many other parts of the refinery where heat recovery is an issue. For the following discussion, the crude preheat train is used as an example, but similar calculations can be made for any heat-recovery duty where heater savings are achieved. For detailed calculations, please see Appendix A at the end of the article.

For a 100 000 bpd refinery, around 120 MW of energy input is needed to preheat the crude from 25–350°C. Most of this energy is recovered, either from the atmospheric residue or from other hot fractions in the refinery. The final part of the preheating is carried out in a fired heater.

It is obvious that the more energy that can be sourced via heat recovery and implemented prior to the heater, the greater the savings in heater size, and the greater the reductions in energy consumption and emissions from that heater. For 1.0 MW less energy input in the heater, around 1070 tons of fuel is saved. Using a fuel cost of $40/bbl (the

***Figure 3** Difference in heat-transfer area and ground space for a S&T compared with a CPHE installation in a heat-recovery installation (atmospheric residue: 288 $\rightarrow$162°C; crude:107 $\rightarrow$ 201°C; heat load: 7MW). S&T total heat transfer area is more than 1000 m^2 and CPHE heat transfer area area is ony 170 m^2*

Figure 4 *CPHE provides unrestricted access to the heat-transfer surface for mechanical cleaning*

Annual savings for 1 MW of increased heat recovery, resulting in a 1 MW reduction in energy needs for heater

Savings	Ton/year and MW	$/year and MW
Heater fuel	1070	300 000
CO_2 emissions	2500	50 000
SO_x and NO_x emissions	30	30 000

Major assumptions: fuel cost $40/bbl, 65% heater efficiency, 0.6% S and 0.25% N in fuel, CO_2 credit $20/ton and SO_x and NO_x credits $1000/ton

Table 3

normal level for internal energy charges), a substantial amount of money, amounting to as much as $300 000, can be saved for every year of operation.

Substantial savings can also be made on emissions. For each ton of natural gas burned, approximately 2.6 tons of CO_2 are released. According to the terms of the Kyoto Protocol, one CO_2 emission allowance (also called a permit or credit) gives the right to emit one metric ton of CO_2 (or CO_2 equivalent). Companies can buy and sell CO_2 credits, and market forces will then set the price of the traded credits. Some US emission traders have forecast market prices of around $20 per credit before 2008 (the first commitment year of the Kyoto Protocol). This is approximately half of the shortfall fines (€40 per credit) proposed in Europe.[1] Using $20 per credit, an additional $50 000 can be saved on reduced CO_2 emissions for each 1.0 MW of energy reduction in the heater on an annual basis.

Other emissions that can be traded are SO_x and NO_x permits. Refiners with emissions below the permitted cap values can sell their surplus permits, and vice versa. The normal trading price for SO_x and NO_x is in the region of $1000 per ton, and for each ton of fuel burned about 15kg of SO_x (assuming 0.6% S in the fuel) and 12kg of NO_x (including both thermal and fuel NO_x and assuming 0.25% N in the fuel) are released. This means that an additional $30 000 can be saved on reduced SO_x and NO_x emissions for each 1.0 MW of energy reduction in the heater on an annual basis (Table 3).

CPHE heat-recovery cases

Crude preheating

A refinery in the US wanted to increase the preheating of its crude. Thermal and hydraulic constraints in the S&Ts had previously limited the preheating to 239°C prior to the heater. It proved possible to achieve additional preheating by installing a CPHE, operating as:

HVGO	298 -> 271°C
Crude	263 <- 239°C

Q = 7.0 MW

Note that all the payback calculations listed in this discussion are based on current market prices and monetary values. Savings are calculated as laid out in Appendix A.

The installed cost (investment plus installation) of two CPHEs (one on standby) was around $1.1 million (Figure 5), but the annual savings in fuel costs and emission reductions amounted to more than $2.7 million ($2.2 million in fuel savings and $500 000 in emission reductions). This energy-saving project thus resulted in a payback time of less than five months. The CPHEs have been in operation since 2004.[2]

A refinery in Switzerland (Figure 6) wanted to use its virgin naphtha overhead vapour from the atmospheric distillation tower to preheat both crude and boiler feed water (BFW). Important criteria when selecting the correct equipment were compactness (as the installation was high above the ground) and corrosion resistance (as the virgin naphtha contained a high level of chlorides). The refinery therefore chose to go for four CPHEs made of titanium (two in parallel and two in series), operating as follows:

Step 1 - BFW preheating:

Naphtha	124 -> 101°C
BFW	95°C <-

Step 2 - Crude preheating:

Naphtha	101 -> 77°C
Crude	71 <- 25°C

Q = 16.5 MW.

Each CPHE pair requires an installation space of about 15m² (including the service area) and weighs around 25 tons.

The energy recovered corresponds to annual fuel and emission savings of $5 million and $1.3 million respectively. With an installed cost of $2.2 million, the payback time was little more than four months. The two CPHE pairs have been in successful operation since 1997.

BFW preheating

A refinery in the UK was looking at heat recovery from the atmospheric residue to preheat its BFW. Two possible setups were discussed: one a conventional solution based on traditional S&T features, and one featuring increased heat recovery, as shown in Table 4.

The conventional approach required

Figure 5 *Two CPHEs used for crude preheating by means of HVGO*

Figure 6 *One of the two CPHE pairs mounted on top of each other for preheating BFW and crude by means of naphtha toppings*

Figure 7 *Heat-recovery installation in UK, preheating BFW by means of atmospheric residue*

Conventional vs increased heat recovery for BFW preheating

		Conventional case	Increased heat recovery
Atm res	300 t/h	165 ->103°C; NTU = 1.7	165 -> 90°C; NTU = 3.4
BFW	140 t/h	132 <- 60°C ; NTU = 2.0	148 <- 60°C; NTU = 4.0
		Q = 12 MW	Q = 14.5 MW

Table 4

two S&Ts in series or one CPHE, while the solution with increased heat recovery required four S&Ts in series or two CPHEs in parallel, rendering the S&T solution not feasible. However, as the CPHE solution meant increased fuel and emission savings of $750 000/year and $200 000/year respectively, the refinery chose to go for the increased heat-recovery option with CPHE equipment. The additional CPHE increased the installed cost by $300 000, but with the increased heat recovery the payback of this additional investment was less than four months. The CPHEs have been in operation since 2002 (Figure 7).

A refinery in Canada was going to install a new hydrotreater, with a required heat input of 4.0 MW for the production of low-sulphur gasoline. Since reduced energy consumption and lower CO_2 emissions were essential, a new furnace was ruled out and the energy was instead to be added as steam. This additional steam was to be generated by recovering heat to the BFW from the FCC fractionator overhead vapours. At the same time, the refinery chose to maximise heat recovery from these vapours, recovering a total of 14 MW. The existing S&T coolers could not handle this heat-recovery requirement, and the refinery therefore selected eight CPHEs with Hastelloy plates (four in parallel and two in series), operating as follows:

Step 1 - BFW preheating:

Overhead vapour	141 -> 87°C
BFW	128 <- 43°C

Q = 14 MW.

Step 2 - cooling:

Overhead vapour	87 -> 29°C
Cooling water	45 <- 24°C

Q = 9 MW.

The heat recovery of 14 MW means annual savings in fuel and emissions amounting to $4.3 million and $1.1 million respectively, and with the installed cost of around $7 million the installation was paid back in 15 months. The CPHEs have been in operation since 2002 (Figure 8).

Steam generation

A refinery in The Netherlands wanted to use mid pump-around (MPA) from the hydrocracker fractionator to generate steam. Due to the close temperature program, it would not have been feasible to do this using S&Ts. Instead, four CPHEs were used: one to preheat the BFW, two for the actual steam generation and one for superheating the steam, as follows:

Step 1 - BFW preheating:

MPA	186 -> 181°C
BFW	179 <- 120°C

Step 2 - Steam generation:

MPA	220 -> 186°C
Water/steam	180 <- 179°C

Step 3 - Steam superheating:

MPA	221 -> 220°C
Steam	215 <- 180°C

Q = 8.4 MW.

With this arrangement, 8.4 MW of heat is recovered from the MPA and 15 tph of 9.0 barg superheated steam is generated. This means annual savings in fuel and emissions amounting to $2.5 million and $600 000 respectively. With an installed cost of approximately $800 000, the installation was paid back in around three months. Or, using a price of $10/ton of 9.0 barg steam, the value of the generated steam amounted to $1.2 million. When combined with the emission savings, this gives a project payback time of only five months. The CPHEs have been in service since autumn 2002.

Heat recovery in desalting units

In the desalting process, the desalter water from the crude oil desalter is normally used to preheat the fresh feed water going to the desalter. As much heat recovery as possible is required, but the desalter water contains substantial quantities of chloride. This means carbon steel is often not sufficiently corrosion resistant. A solution involving large S&Ts made of stainless steel or another high-performance alloy material would be extremely expensive.

Due to corrosion problems, a refinery in Germany needed to replace its six carbon-steel S&Ts used for heat recovery in the desalting process. Due to the need for corrosion-resistant material, the refinery decided on three CPHEs featuring Hastelloy plates, installed in series and operating as follows:

Figure 8 CPHE installation for recovery of a total of 14 MW from FCC overhead vapours into BFW loop.

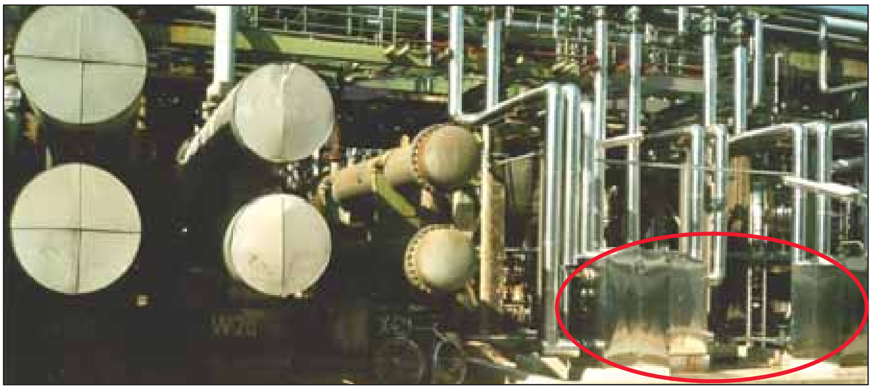

Figure 9 CPHE installation for heat recovery in the desalting process. The three CPHEs marked with a circle replace the six S&Ts to the left

Step 1 - Heat recovery:

Desalted water	130 -> 50°C
Fresh water	120 <- 40°C

Step 2 - Cooling:

Desalted water	50 -> 30°C
Cooling water	40 <- 25°C

Q = 3.7 MW.

As can be seen in Figure 9, the refinery not only upgraded its heat exchangers to a corrosion-resistant material, it also substantially reduced the size of the installation (the three CPHEs marked with a circle replaced the six S&Ts to the left, as shown in Figure 9). The heat recovery achieved corresponds to energy and emission savings of around $1.1 million and $300 000 respectively. With an installed cost of approximately $600 000, this meant the installation was paid back in around five months. The CPHEs have been in operation since 1997.

Conclusion

It is a well-known fact that compact CPHEs, with their improved turbulence and counter-current flow, can achieve much higher heat-transfer efficiencies than traditional S&Ts, thereby reducing the heat-transfer area needed.

This is especially important in heat recovery, where the use of CPHEs makes it possible to carry out demanding energy-recovery duties that in some cases would not even be feasible using S&Ts, or would require many large S&Ts connected in series and thus suffer from hydraulic constraints and fouling problems.

The increased heat recovery means substantial savings, both in terms of fuel savings and savings on emissions from heaters and boilers. With today's rocketing energy prices, the Kyoto Protocol and acid rain, these are facts that become ever more important when calculating the payback for a project.

Since CPHEs provide compact, lightweight solutions with minimal installation footprint and service ground space needs, the installation cost is normally 33–75% lower than for bulky S&Ts. Payback times of less than six months, including installation cost, are often feasible when considering heat-recovery projects that feature CPHEs.

Other advantages with using CPHEs include the reduced heat-transfer area, which makes it possible to utilise materials that are highly corrosion-resistant, and the low hold-up volume, which enables the unit to respond more quickly to any changes in the process operating parameters, making it easier to start up and shut down the process.

From a service point of view, the highly turbulent flow through the heat exchanger channels, measured as wall shear stress, ensures the heat exchanger is kept clean, resulting in longer service intervals. When cleaning is needed, the unrestricted access to the heat-transfer surface reduces the downtime and maintenance efforts to a minimum. If chemical cleaning can be carried out, chemical consumption and disposal costs are reduced on account of the much lower hold-up volume.

Many refineries have started to see the huge benefits of using CPHEs in their processes. Most of the installations are for heat-recovery applications (crude preheating, feed/effluent heat recovery and BFW preheating) when installation space or weight is crucial (overhead condensers) and exotic materials are needed to deal with corrosion issues (desalter water, naphtha toppings and alkylation mix). There is no lack of testimonials from operators who are convinced of the benefits.

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3 Takacs T J, Juedes D L, Crane I D, Method estimates NO_x from combustion equipment, *Oil & Gas Journal*, 21 June 2004.

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APPENDIX A

Detailed energy and emission savings calculations

Crude preheating

$$Q = m \times Cp \times (T_2 - T_1) = \frac{100\,000}{7.2 \times 24 \times 3.6} \times 2.3 \times (350 - 25) = 120\,161\ kW = 120\ MW$$

Q = Heat load in kW
m = Flow rate of crude in kg/s
Specific volume for crude = 7.2 bbl/ton
Cp = Specific heat for crude = 2.3 kJ/kg
T_1 = Crude start temperature = 25°C
T_2 = Crude final temperature = 350°C

Fuel savings in heater

Calculations based on natural gas (NG) as heater fuel:

$$m_{NG} = \frac{t}{H_vNG \times n_{eff}} = \frac{3600 \times 24 \times 350}{11\,500 \times 4.19 \times 0.65} = 966\ ton/year\ and\ MW$$

m_{NG} = NG savings in heater in ton/year and MW
t = Time in seconds/year (assuming 350 operating days/year)
H_vNG = Latent heat for NG = 11 500 tcal/ton of natural gas
Conversion factor from tcal/ton to MW/ton = 4.19
n_{eff} = Radiant cell efficiency of heater = 65%

$$price_{NG} = \frac{H_vNG}{H_vFuel} \times price_{fuel} = \frac{11\,500}{10\,400} \times 40 \times 7.2 = \$318/ton$$

$Price_{NG}$ = Price for NG, assuming same MW cost as fuel, in $/ton
H_vNG = Latent heat for NG = 11 500 tcal/ton
H_vFuel = Latent heat for fuel = 10 400 tcal/ton
$Price_{Fuel}$ = Internal energy charges for fuel = $40/bbl
Specific volume for fuel = 7.2 bbl/ton

$$savings = m_{NG} \times price_{NG} = 966 \times 318 = \$307\,634/year\ and\ MW$$

CO_2 emission savings in heater

Calculations based on natural gas (NG) as heater fuel:

$$m_{CO2} = m_{NG} \times wt\%_C \times \frac{Mw_{CO2}}{Mw_C} = 966 \times 0.71 \times \frac{44}{12} = 2514\ ton/year\ and\ MW$$

m_{CO2} = CO_2 emission reductions in heater, in ton/year and MW
m_{NG} = NG savings in heater in ton/year and MW
$wt\%_C$ = Carbon content in natural gas = 71 wt%
Mw_{CO2} = Molecular weight of CO_2 = 44 g/mole
Mw_C = Molecular weight of C = 12 g/mole

$$savings = m_{CO2} \times price_{CO2} = 2514 \times 20 = 50\,296\ \$/year\ and\ MW$$

$price_{CO2}$ = estimated CO_2 credit price = $20/ton

- SO_x emission savings in heater

Calculations based on heater fuel containing approx. 0.6% S

$$m_{fuel} = \frac{t}{H_vfuel \times n_{eff}} = \frac{3600 \times 24 \times 350}{10\,000 \times 4.19 \times 0.65} = 1100\ ton/year\ and\ MW$$

m_{fuel} = Fuel savings in heater in ton/year and MW
t = Time in seconds/year (assuming 350 operating days/year)
$H_v$fuel = Average latent heat for fuel = 10 000 tcal/ton of fuel
Conversion factor from tcal/ton to MW/ton = 4.19
n_{eff} = Radiant cell efficiency of heater = 65%

$$m_{SOx} = m_{fuel} \times wt\%_S \times \frac{Mw_{SOx}}{Mw_S} = 1110 \times 0.006 \times \frac{80}{32} = 16.7\ ton/year\ and\ MW$$

m_{SOx} = SO_x emission reductions in heater, in ton/year and MW
m_{fuel} = Fuel savings in heater in ton/year and MW
$wt\%_S$ = Sulphur content in fuel = 0.6 wt%
Mw_{SOx} = Molecular weight of SO_3 = 80 g/mole
Mw_S = Molecular weight of S = 32 g/mole

$$savings = M_{SOx} \times price_{SOx} = 16.7 \times 1000 = \$16\,655/year\ and\ MW$$

$Price_{SOx}$ = estimated SO_x credit price = $1000/ton

- NO_x emission savings in heater

The total figure for NO_x emissions has two main sources: thermal NO_x (high temperature oxidation of molecular N_2 in air) and fuel NO_x (direct oxidation of organic nitrogen in fuel). These are calculated separately and then added together to give the total NO_x emission savings.

o Fuel NO_x

Calculations based on heater fuel containing approx. 0.25% N:

$$m_{fuel} = \frac{t}{H_vfuel \times n_{eff}} = \frac{3600 \times 24 \times 350}{10\,000 \times 4.19 \times 0.65} = 1110\ ton/year\ and\ MW$$

m_{fuel} = Fuel savings in heater in ton/year and MW
t = Time in seconds/year (assuming 350 operating days/year)
$H_v$fuel = Average latent heat for fuel = 10 000 tcal/ton of fuel
Conversion factor from tcal/ton to MW/ton = 4.19
n_{eff} = Radiant cell efficiency of heater = 65%

$$m_{NOx} = m_{fuel} \times wt\%_N \times \frac{Mw_{NOx}}{Mw_N} = 1110 \times 0.0025 \times \frac{46}{14} = 9.12\ ton/year\ and\ MW$$

m_{NOx} = NO_x emission reductions in heater, in ton/year and MW
m_{fuel} = Fuel savings in heater in ton/year and MW
$wt\%_N$ = Nitrogen content in fuel = 0.25 wt%
Mw_{NOx} = Molecular weight of NO_2 = 46 g/mole
Mw_N = Molecular weight of N = 14 g/mole

o Thermal NO_x

Thermal NO_x is primarily a function of flame temperature, which in turn is a function of the fuel latent heat, stoichiometric air/fuel ratio, excess air, air preheat, flue gas recirculation, humidity, low-NO_x burners, burner intensity, heat-removal rate and gas-mix residence time.[1] ExxonMobil R&D reports a comprehensive calculation method for thermal NO_x.[3]

The following equation is given for a modern heater, operating 350 days/year with about 150°C air preheating:

$$m_{NOx} = 3 \pm 1.3\ ton/year\ and\ MW$$

The user may choose to estimate small variations on either side of the base number if some factor that will affect the flame temperature is known to be present (for example, a high hydrogen content in fuel gas will raise the flame temperature and hence the thermal NO_x, while the absence of air preheating will lower these).

$$Totalm_{NOx} = 9.12 + 3 = 12.1\ ton/year\ and\ MW$$

$$savings = m_{NOx} \times price_{NOx} = 12.1 \times 1000 = \$12\,120/year\ and\ MW$$

$Price_{NOx}$ = estimated NO_x credit price = $1000/ton

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