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Fired heater design and operation to minimize coking

Application in refinery residue services

Carlos Jorge Batista Alcobia

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Presidente:	Prof. João Carlos Moura Bordado
Orientador:	Prof.ª Maria Filipa Gomes Ribeiro
Co-Orientador:	Eng.ª Marie-Andrée Senegas
Vogal:	Eng. Philippe Dolbecq

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I will always remember this training which was my first contact with work in the process industry. I hope my future career will be shared with such kind and supporting people as the ones I worked with at TOTAL, especially everyone in the DAT team.

Abstract

English

The main objective of this work was to specify basic fired heater design and operation criteria that help minimize the formation of coke in vacuum distillation unit fired heaters. Additionally, two heaters with severe coking problems, one from a vacuum distillation unit and another from a visbreaking unit, were analyzed in order to troubleshoot them.

This work presents design specifications, operational advice and revamp considerations on vacuum distillation unit fired heaters. Vacuum heater operation data from TOTAL refineries was retrieved and compared to design guidelines found in literature. Fired heater simulation software was used to model several heaters and to study the influence of several parameters on the coke formation rate. The two heaters with coking problems were also modelled to search for possible design or operation problems and latter to analyze possible problem scenarios.

The main conclusion is that the most important design parameters in minimizing fired heater coke formation are mass velocity and fluid film temperature in the tubes. The recommended mass velocity for fired heaters in vacuum distillation service is 1500 $kg/m^2.s$ at the radiant section inlet. The fluid film temperature should be kept below 460 ºC in all tubes except the heater's outlet tube.

The Donges visbreaker fired heater was found to suffer from high heat flux to the shock tubes and possibly from charge quality variation problems. The Leuna vacuum distillation unit fired heater is probably affected by hotspots caused by high refractory re-radiation from the firebox's roof.

Key words

COKING, COKE FORMATION, FIRED HEATER, VACUUM HEATER, VISBREAKING

Português

Este trabalho teve como principal objectivo especificar critérios básicos de dimensionamento e operação de fornos, que minimizem a formação de coque em fornos de unidades de destilação de vácuo. Dois fornos com problemas severos de formação de coque, um de uma unidade de destilação de vácuo e outro de uma unidade de visbreaking, foram analisados numa tentativa de minimizar o problema.

Este relatório recomenda parâmetros de dimensionamento e conselhos de operação e revamp de fornos de unidades de destilação de vácuo. Foi recolhida informação de fornos das unidades de destilação de vácuo da TOTAL e comparada aos parâmetros de dimensionamento encontrados na literatura. Software de simulação foi usado para modelar vários fornos e estudar a influência de diversos parâmetros na taxa de formação de coque. Os dois fornos com problemas de formação de coque foram também modelados para procurar possíveis problemas de dimensionamento ou de operação.

A principal conclusão é que os parâmetros de dimensionamento mais importantes na minimização da formação de coque são a velocidade mássica e a temperatura de filme nos tubos. O valor recomendado para a velocidade mássica à entrada da zona de radiação é 1500 $kg/m^2.s^{-1}$ e a temperatura de filme deve ser mantida abaixo dos 460 ºC em todos os tubos excepto o de saída do forno.

Concluiu-se ainda que ambos os fornos com problemas sofrem de elevado fluxo de calor por re-radiação do refractário, embora em zonas diferentes. O forno do visbreaker de Donges possivelmente também terá problemas de variação da qualidade da carga.

Palavras-chave

COQUE, FORMAÇÃO DE COQUE, FORNO, DESTILAÇÃO DE VÁCUO, VISBREAKING

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List of symbols and abbreviations

Cp: specific heat capacity in J/(kg.K)
CReG: Centre de Recherche de Gonfreville
CReS: Centre de Recherche de Solaize
D: diameter in meters
DAT: Département Assistance Technique
F: factor accounting for heat flux variations
FGR: fuel gas recirculation
FRNC-5: fired heater simulation software produced by PFR Engineering.
h: heat transfer coefficient in W/(m^2.K)
k: thermal conductivity in W/(m.K)
LHV: lower heating value of fuel
NOx: name as nitrogen monoxide and dioxide are collectively referred to
P: pressure
Pr: Prandtl number
Q_{mA}: mass velocity in kg/(m^2.s)
R: ratio of critical velocity in gas to critical velocity in liquid (0.333 is a good approximation for most refinery hydrocarbon streams)
Re: Reynolds number
SCR: selective catalytic reduction
SNCR: selective non-catalytic reduction
T: temperature
v: velocity
VDU: vacuum distillation unit. Distillation unit operating at pressures below 1 atm used to recover hydrocarbons from the atmospheric distillation residue and therefore minimize the amount of residue left from the initial crude.
VGO: vacuum gas oil. Distillate obtained from vacuum distillation of crude atmospheric residue.
w: mass fraction
x: vapor phase mass fraction
Xfh: fired heater simulation software produced by HTRI.
α: sonic velocity
μ: absolute viscosity in Pa.s
ρ: density in kg/m^3
ϕ: heat flux in W/m^2

Subscripts

2p: two-phase
ave: average
b: fluid bulk
c: critical
cir: circumferential
conv: convection
film: oil film adjacent to the tube wall
g: gas
i: inside tube
l: liquid
L: longitudinal
rad: radiation
v: vapor
w: tube wall

1. Introduction

Coke is a solid substance with a very high carbon content that deposits on the walls of refining equipment and piping. In fired heaters this introduces an additional resistance to heat transfer, raising tube metal temperatures and forcing equipment stops for decoking. When the tube metal temperature approaches the metallurgic limit the heater must be decoked or the tube will suffer irreparable damages and may burst, forcing an emergency stop and even risking an explosion. Besides raising the metal temperature of the tubes, coke also increases pressure drop but this isn't usually limiting to heater operation.

It is important to note that coke by itself does not pose a problem to fired heater operation. The problem is the metal temperature of the tubes that is raised by coke deposited in the tubes. When tube metal temperature approaches the tube metallurgic limit the heater must be decoked or the tube will suffer irreparable damages and may burst, forcing an emergency stop with extended stop for repairs and even risking an explosion.

The formation of coke deposits in refinery residue service fired heaters increases tube metal temperatures restricting unit run length and in some cases causing tube failure. With today's focus on processing heavier crude and improving VGO recovery, the severity in fired heaters has been increased aggravating the coking problem. This highlights the need for better fired heater design. Additionally, a well design fired heater can still present coking problems if it is not operated correctly and operating costs are always a concern. The significant impact of heater operation on coking calls for better understanding of its influence.

It was the aim of this work to investigate and propose good fired heater design practices in vacuum distillation unit service and also to discuss and propose correct fired heater operation practices.

Available information in the form of scientific articles, internal TOTAL documents and technical courses on fired heaters was investigated in order to find out which are the most important parameters to specify. A lot of information was found on fired heater design, operation and revamp. Some of this information was contradictory, other considered insufficient and in the case of design specifications, different sources sometimes propose different values. This data alone were considered insufficient to allow any conclusions as to which design values were the best.

The need for clarity and confirmation led me to analyze refinery data, simulate several fired heaters using specialized software and to arrange a meeting with an engineer from a leading fired heater engineering company to discuss fired heater design specifications. A more thorough investigation of some of the design specifications influence on coking was made, mostly through fired heater simulation software.

This work presents design specifications, operational advice and revamp considerations on vacuum distillation unit fired heaters and their importance and influence is also discussed.

Today's cost of energy and the increasing concern with the environment has made energy conservation an important issue. Old heaters have low energy efficiency and revamps are in many cases necessary to lower costs and meet new environmental regulations. Several revamp options are discussed in this work, as well as their advantages and disadvantages.

The Leuna refinery VDU heater has been in operation since 1997 and has suffered from shorter than expected run lengths due to high skin temperatures at various locations in the radiant coils. The Donges refinery visbreaker heater suffered an accident with the rupture of a shock tube and has been plagued by rapidly rising skin temperatures in the shock tubes ever since, severely limiting run length. I was invited to participate in the troubleshoot of both heaters, analyzing the problems through software simulations and by looking at refinery data with the help of previous reports on the problems. These analyses helped finding the likely causes and solutions were proposed.

To better understand the information presented in this report it is perhaps best to make a small introduction to fired heaters and vacuum distillation units. An explanation of what is coke and on how coking occurs is also provided in the following text.

1.1. Fired heaters

API Standard 560 [1] defines a fired heater as "an exchanger that transfers heat from the combustion of fuel to fluids contained in tubular coils within an internally insulated enclosure". Most of the heat transfer is made through radiation although convection also plays an important part.

Fired heaters can be divided into the following key areas:

- **Radiant section** or **Firebox** – heat is transferred mostly by radiation in this section
- **Convection section** – heat is transferred mostly by convection in this section
- **Stack** – creates draft in the heater to allow the in-flow of combustion air and the outflow of flue gas
- **Air pre-heater** (optional) – increases heater efficiency by recovering heat from the flue gas

The different fired heater components can be seen in figure 1.1, except for the air pre-heater which is shown in figure 1.2. The names of the components noted in the picture are:

1 - Access door
2 - Arch
3 - Breeching
4 - Bridgewall
5 - Burner
6 - Casing
7 - Convection section
8 - Corbel
9 - Crossover
10 - Tubes
11 - Extended surface
12 - Return bend
13 - Header box
14 - Radiant section
15 - Shock or shield section
16 - Observation door
17 - Tube support
18 - Refractory lining
19 - Tube sheet
20 - Pier
21 - Stack
22 - Platform

In the radiant and convection sections, where heat transfer takes place, we can define a hot and a cold side like in any other heat exchanger:

- **Tubeside** or **Process side**, where the cold fluid circulates, in this case an heavy oil
- **Fireside**, where the hot fluid circulates, which is combustion gas, also called flue gas

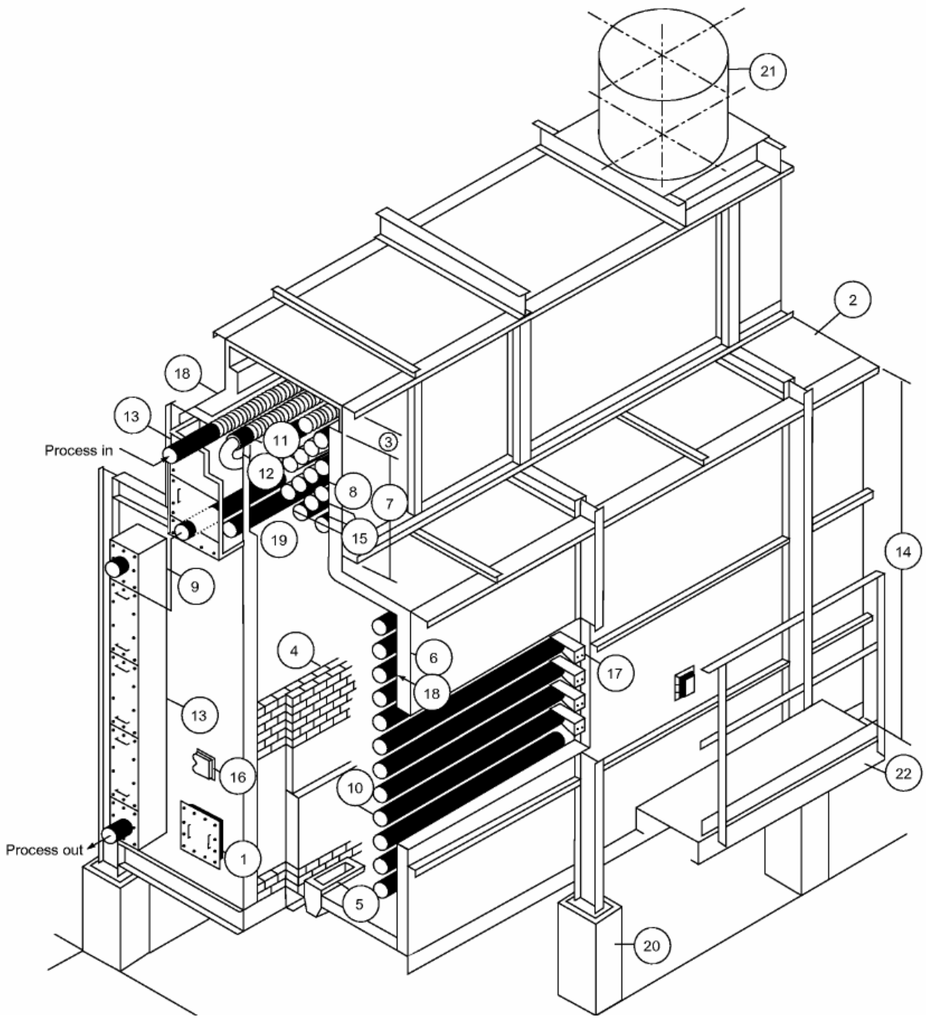


Figure 1.1 - Fired heater components [1]

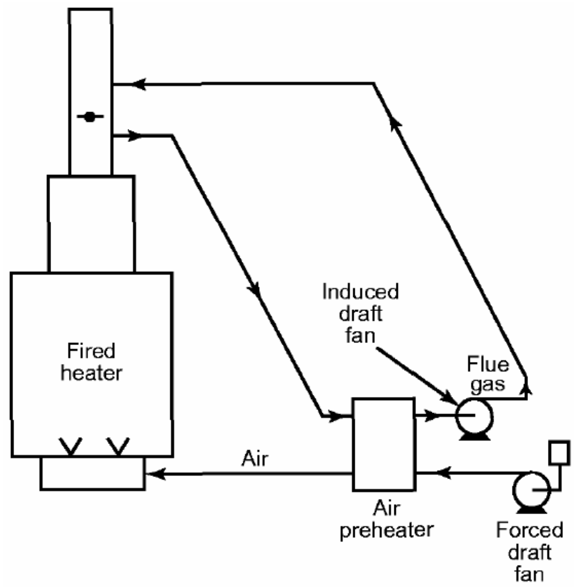


Figure 1.2 - Air pre-heat system [1]

Heat is provided by burning of a fuel which can be fuel oil, fuel gas or a mixture of both. The heat rate released by the combustion is called firing duty. Of the firing duty about 15% is lost, most of it in the flue gas released to the atmosphere. Around 1.5% to 2% of the firing duty is lost through the walls of the heater.

Because the flue gas contains SO_3, which can condensate on cold metal surfaces and corrode the heater's structure, the stack temperature is kept at least 50 ℃ above the acid mist dew point temperature. These heat losses in the flue gas are inevitable and therefore the maximum thermal efficiency achievable in a fired heater is 92% of the fuel's Lower Heating Value – LHV. The rest of the heat is transferred to the process fluid inside the tubes and is called heat duty.

Due to the cost of fuel and environmental concerns, nowadays heaters are designed to be as thermal efficient as possible. A convection section allows the design engineer to lower the flue gas temperature to around 300-350 ºC depending on the size of the convection section, the oil inlet temperature and on the use of a steam coil. Every new heater is equipped with a convection section in order to recover as much heat as possible. When used, between 30% and 35% of the heat duty is transferred in the convection section and the remaining 65% to 70% in the radiant section of the heater. This will reduce the size of the radiant section as well as the fuel consumption. In existing heaters it can be used to increase capacity or to lower radiant section heat flux.

Further heat recovery can be achieved with an air pre-heater. This is a heat exchanger that transfers heat from the flue gas that exits the convection section to the combustion air fed to the burners. This can save fuel and the limit of the heat recovery is the acid mist dew point temperature described above. This is usually used only on large heaters because it requires a forced draft system, therefore the cost considerations must include its cost. But only with an air pre-heater can we achieve maximum heater efficiency. More on air pre-heaters will be discussed in section 2.10.2.

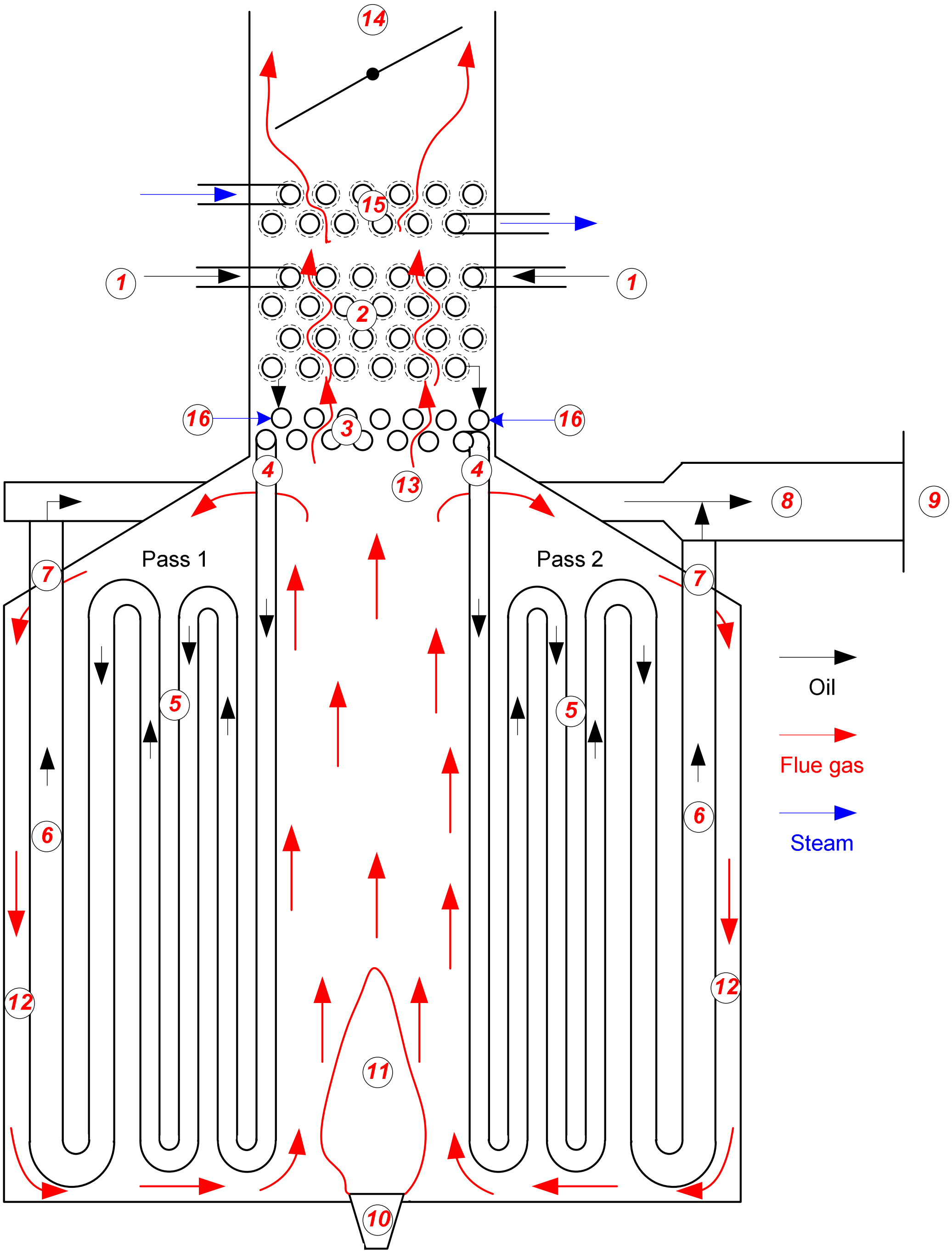


Figure 1.3 - Diagram of a fired heater

A drawing of a fired heater is presented in figure 1.3 with arrows that illustrate the flow of oil, steam and flue gas. The points marked in the diagram are described below with typical values in vacuum distillation service. Most values are presented in the form of an interval that covers the usual range in vacuum heaters. Heater thermal efficiency is 75% - 90% or 85% – 92% for recent and revamped heaters. Heater heat duty, the heat transferred to the charge, can go from 8 MW up to 60 MW in one fired heater. If more duty is necessary usually two or more fired heaters are used, in the Antwerp refinery two fired heaters and a pre-heater provide 87 MW.

1 - **Heater inlet** – T: 260 – 360 ºC; P: 3.5 – 9 atm abs; feed rate: 60 – 650 ton/h; feed specific gravity @ 15 ºC: 0.90 – 1.03
2 - **Convection section tubes with extended surface** – fin or studded tubes; 7 to 11 rows; 2 to 8 tube passes
3 - **Shock tubes** – naked tubes; 2 or 3 rows subject to significant radiation and convective heat transfer which results in high average heat flux: 40 – 50 kW/m^2, can be as high as 80 kW/m^2
4 - **Radiant section inlet** – T: 320 – 370 ºC
5 - **Radiant section tube coil** – average heat flux: 20 – 40 kW/m^2; 2 to 8 tube passes; 3 to 12 skin thermocouples per pass; limiting metal temperature: 650 – 705 ºC depending on alloy; residence time 50 – 60 s; maximum film temperature (with the exception of the outlet tube) is usually below 460 ºC but may rise up to 485 ºC in some heaters
6 - **Heater outlet tube** – film temperature can pass 600 ºC
7 - **Heater outlet** – T: 390 – 430 ºC; P: 0.3 - 0.7 bar abs; vapor weight fraction: 20% – 50%; two-phase flow regime: annular or mist; fluid velocity can be 55% to 95% of the critical velocity (calculated with FRNC-5 for several heaters)
8 - **Transfer line** – temperature drop: 5 – 20 ºC, can be as high as 40 ºC; pressure drop: 0.2 - 0.5 bar, can be as high as 1 bar; fluid velocity can reach the critical velocity
9 - **Distillation column flash zone** – T: 380 – 410 ºC; P: 60 – 130 mbar; vapor weight fraction: 50% – 75%; VGO cut-point: 500 – 600 ºC
10 - **Flame** – radiating temperature: 1650 - 1950 ºC; flame height: 1/3 to 2/3 of the heater height, oil flames are larger than gas flames
11 - **Hot flue gas** – flue gas temperature is much lower than flame temperature due to mixing with the recirculated cold flue gas. Flue gas also cools as heat is transferred to the tubes while it crosses the firebox
12 - **Cold flue gas recirculation** – recirculated flue gas rate can be 2 to 4 times the flue gas rate produced by the flames; cold flue gas temperature is near the bridgewall temperature of the heater
13 - **Arch** – T Bridgewall: 700 – 900 ºC; P: -0.05 – -0.15 inH_2O; excess O_2: 2% - 6%;
14 - **Stack** – T flue gas: 240 – 315 ºC or 370 – 420 ºC if air pre-heating is not used
15 - **Steam generator or superheater** – low or medium pressure superheated steam can be produced from condensate or saturated steam to recover heat from flue gas

16 - **Coil steam injection** – coil steam is usually injected at a rate of 0.2% to 0.5% weight of the charge; usual injection points are the shock bank inlet and the radiant section inlet

In the annex section, table A.1 presents a collection of values from TOTAL VDU units. Mr. Berman has written a good collection of articles on fired heaters [2] and a summary of heater technology is also available at TOTAL [3].

1.2. VDU – Vacuum distillation unit

The specific type of fired heater under consideration in this work is part of the VDU. The VDU consists not only of a fired heater but also of a vacuum distillation column and a transfer line, which connects the other two equipments.

Vacuum units have acquired extra significance in the past two decades due to the fact that crude oils are becoming heavier. There is a continued requirement for heavy distillate feedstock which places much emphasis on the need to cut deeper into atmospheric residue.

When designing or revamping a VDU, the system must be considered as a whole. Because it is an integrated system, design considerations in one of the equipments affect the others. For example, in order to meet the distillate separation targets in the column, a certain amount of heat must be input in the heater.

The heater is used to provide sensible and latent heat to the feed, so that it provides the column with the necessary enthalpy for the separation. The flash zone of the vacuum distillation column must operate under certain conditions of pressure, vapor fraction and temperature in order to meet the desired VGO yield.

The transfer line is designed for reasonable pressure drop based on mechanical flexibility and spatial considerations due to the locations of the column and the heater. The pressure and temperature drop in the transfer line will determine the temperature and pressure in the heater's outlet. The temperature drop in the transfer line is essentially due to isenthalpic expansion occurring with pressure drop.

Finally, the heater must be considered having in mind the conditions at its outlet and not at the inlet [4] [5] [6] [7]. The pressure, temperature and vapor fraction profiles of the VDU extend from the column flash zone to the heater inlet. These are the most important design variables because they influence everything, from the coking rate and pumping costs to the size of the heater.

Therefore, **when designing or revamping the heater for a VDU, one must have the column flash zone operating variables as the target, not the heater's**.

1.3. Coking

Coke is a solid substance with a very high carbon content that deposits on the walls of refining equipment and piping. In fired heaters this introduces an additional resistance to heat transfer, raising tube metal temperatures and forcing equipment stops for decoking. When the tube metal temperature approaches the metallurgic limit the heater must be decoked or the tube will suffer irreparable damages and may burst, forcing an emergency stop and even risking an explosion. Besides raising the metal temperature of the tubes, coke also increases pressure drop but this isn't usually limiting to heater operation.

Inorganic salts may also exist in the feed and deposit on fired heater tube walls where vaporization occurs, forming scales with high thermal resistance. These salt deposits have higher thermal resistances than petroleum coke.

Coke is a product of the thermal cracking of oils and is linked to the quality of the charge, especially the asphaltene and sodium content. The starting temperature for thermal cracking depends on the thermal stability of the oil in question but is in the range of 370 ℃ to 400 ℃. [8]

The occurrence of thermal cracking does not necessarily imply the formation of coke on the tube walls. The rate of thermal cracking depends on the temperature of the oil and its residence time at the given temperature. On vacuum heaters it can be problematic at temperatures above 420 ℃ but even at higher temperatures the coking rate can be managed if the residence time is low enough. 800 ℉ (427 ℃) is widely considered as the starting limit for thermal cracking, not because there is no thermal cracking below this temperature but because above it thermal cracking will be enough to lead to coking problems.

According to Shell Global Solutions [9], **only when the oil reaches temperatures above 460 ℃ does the coking rate become too high for the conditions usually found in a vacuum heater**.

There is a temperature profile from the radiating surface to the bulk of the oil being heated and the highest oil temperatures are in the stagnant oil film adjacent to the tube wall, so we must know these film temperatures at the wall to evaluate coke formation [10]. Bulk oil temperatures alone are not adequate to predict coking problems.

Minimizing oil residence time and/or film temperatures will reduce coking. The coking rate is linearly dependent on residence time and exponentially dependent on temperature.

Conditions that affect the rate of coke formation are:

- oil mass velocity
- oil film temperature / peak heat flux
- oil residence time
- charge asphaltene content
- charge sodium content
- charge oxidation and acidity

Some ways of calculating the coking rate and the coking tendency of oils exist. Studies on the subject were found in scientific papers [8] [11] [12] [13] [14] [15] [16] [17] [18] [19]. However it should be researched if their application is adequate to industrial scale equipment. Some of these methods require data measures taken on a laboratory to predict coke formation tendencies or oil thermal stability. Others rely on historic data of coke deposition with measures of coke thickness and thermal resistance to calculate a coking rate. Since these data are not usually available they would have to be obtained, which is a difficult and lengthy process.

The formation of coke is a complex subject and stating that it starts depositing at a certain temperature due to thermal cracking is a gross simplification. In fact residues have an induction period before coke starts to form. After the induction period, coke forms at a maximum rate and the coking rate decreases.

Mr. Wiehe [20] proposes a model where the heptane soluble substances in the residue inhibit the formation of coke by the asphaltenes. The heptane solubles vaporize or suffer thermal cracking and transform into asphaltenes. When the asphaltene solubility is reached, a phase separation occurs leaving the asphaltenes to form their own phase where they crack very rapidly into coke. Since the rate of asphaltene formation is slower than the rate of coke formation, the formation of coke reaches a maximum just after the induction period due to the asphaltenes already present in the charge and then slows down. **Residues with high asphaltene content will produce more coke as will residues with little asphaltene solubility**.

There are several articles that discuss the problem of asphaltene insolubility and propose different ways to measure charge stability based on solubility maps [8] [14] [21] [22] [23] and on the peptization state of the residue [23]. Also two patents describe a process for blending potentially incompatibly oils: **US 5.871.634** [24] and **US 5.997.723** [25].

There are other charge characteristics that affect the coking potential. **Sodium acts as a thermal cracking catalyst**. When sodium salts are present in residues at concentrations higher than 50 ppm the increase in thermal cracking is significant [19]. Proper desalting of the processed crude is important to limit coking.

Oxidation of the charge is also important. **Oxidized charges usually increase the rate of coke formation**, probably due to polymerization of the oxidized compounds [19]. It may be worth considering blanketing residue tanks with an inert atmosphere if it is not already common practice.

Finally, one other aspect worth mentioning is the quality and thermal resistance of coke. Coke is a designation used for solids constituted mainly of carbon but these have varying percentages of hydrogen and more important, different structures. These different types of coke have thermal conductivities that vary significantly.

The first coke deposited is a porous structure with little structural resistance and filled with fluid. Its conductivity will be similar to the liquid residue’s conductivity which is very low and so film temperature will be high, causing more coke to build up consolidating the deposit. As the deposit becomes less porous by the addition of more coke the thermal conductivity increases, because solid coke is a better conductor than the residue. Time and temperature will degrade this deposit even further decreasing the hydrogen content and hardening it.

Because of this change in coke properties it is not trivial to estimate its thickness from tube wall temperatures. Coke thermal conductivity can vary from the residue’s conductivity, which can be as low as 0.1 W/m.K, to 6 W/m.K for hard coke. Two values of thermal conductivity are referred by several sources, 0.8 W/m.K and around 5 W/m.K. I believe these values can be used for porous and hard coke respectively. But the best way to know the correct conductivity is by collecting samples during decoking stops. TOTAL recommends using a 5.5 or 5.6 W/m.K thermal conductivity, Scarborough [26] 0.3 to 0.9 W/m.K and UOP 4.3 W/m.K for coke and 0.9 W/m.K for inorganic scale from salt precipitation.

The porous coke layer on top of the harder coke will always exist and it is in this porous layer that more coke will continue to form. It is there that liquid is available by diffusion and therefore liquid temperatures are highest. A memo on the aspects of coke formation and its consequences is available at TOTAL [27].

2. Design specifications

The best and most cost effective way of avoiding coking problems in a heater is to properly design it in the first place [28]. Since fired heaters are usually designed by specialized engineering companies, it is the job of the project engineer to provide such companies with minimum design specifications. **Good heater specifications help to ensure proper heater design**.

Since nowadays refinery operation has changed significantly, with heavier crude oils being processed and a focus on increasing run length to maximize profits, so has to change the cost / benefit analysis of vacuum heaters. Many refiners tend to save on the heaters' cost only to end up incurring greater costs with short run lengths and other coking related problems [29].

Hotspots exist in all fired heaters' radiant section; the magnitude depends on the specific heater design. Since many heaters are purchased solely on lowest installed cost, the lowest bidder is forced to supply an "inferior" design. Lowering the cost may include reducing the number of burners, decreasing the distance from the burner center -line to the tube face, increasing the firebox height to width ratio, restricting the number of passes and several other cost-saving considerations. However, many of the techniques that lower cost also cause large heat flux variability throughout the firebox or degrade tubeside operation. Rapid coke formation and high tube metal temperatures are the likely consequences.

It is my belief that a more conservative approach to vacuum heater design will prove worthy in the end by increasing run length and providing room for increases in feed charge. Of course, to increase run length, not only the heater must be working properly but also the rest of the VDU unit.

High localized heat flux causes elevated tube metal temperatures even when no coke is present. This often raises the oil film temperature above the oil's thermal stability and causes coke to form. Even when localized heat flux is not very high, coke can form because of tubeside conditions that result in oil film temperature and residence time above the oil's thermal stability. Mass velocity is an important consideration affecting the rate of coke formation. In summary, localized hotspots can be caused by fireside or tubeside design shortcomings or both.

There are many design considerations that can be analyzed, but they all come down to a cost analysis. It is the job of the design engineer to make a correct cost / benefit analysis while keeping in line with the design specifications provided by the buyer.

The most important design considerations to limit coking are:

- Mass velocity
- Film temperatures
- Coil steam injection

Other design considerations are also discussed, such as:

- Charge quality
- Heat flux
- Type of heater
- Tube coil arrangement
- Outlet tube conditions
- Turndown flow rate
- Convection section
- Air pre-heater
- Draft
- Burners
- Boiling regime
- Two-phase flow regime

A heater more conservatively designed will have less operational problems and it will be much easier to increase the charge, thermal and/or feed rate. This means using:

- High mass velocities
- Low heat fluxes
- Appropriate burner clearance

The investment cost will be higher but this will be offset by the lower operation costs due to:

- Fewer decoking and maintenance stops
- Fewer tubes need to be replaced
- Fewer tube bursting accidents
- More unit availability

The heater should be sized for the worst operation scenario, predicting probable feed composition and rate changes and also changes in vacuum column cut-point targets. TOTAL general heater specifications state that a heater should be able to operate at 120% the rated power for at least one hour. UOP recommends that the heater be sized with a 10% to 15% margin, since a 10% increase in size costs only 5% more and revamps are very expensive.

2.1. Mass velocity

Mass velocity is the most important fired heater design parameter. It sets turbulent flow in single-phase flow; avoids undesirable two-phase flow patterns caused by low flux, such as slug and stratified flow; lowers wall film thickness and increases film heat transfer coefficients; maintains high wall sheer stress which helps removing newly formed coke from the tube walls.

Everyone agrees that setting a minimum mass velocity at the radiant section inlet is a good way to avoid coking problems. What is still open for discussion is the best value for the minimum mass velocity.

Increasing the mass velocity will, in theory, always decrease the coking rate in the tubes since thermal cracking increases linearly with residence time. The problem is that pressure drop also increases linearly with mass velocity. Due to changing two-phase flow regime and decreasing effect in film temperature reduction, there is a limit to the improvement in coking rates achievable by higher mass velocity. Furthermore, since there are other ways of reducing the coking rate, it is a question of economics how much should we increase mass velocity.

2.1.1. Collected values

The values recommended by several sources are shown in table 2.1. Data collected from some vacuum heaters in TOTAL refineries are shown in table 2.2.

Table 2.1 - Recommended minimum mass velocities for vacuum heaters

	Mass velocity kg/m^2.s	
Source	**Inlet tube**	**Outlet tube**
TOTAL	1250	-
Shell	-	293
UOP [30]	1220	293
Heurtey Petrochem	1220	-
PCS [6]	2000	-
Scarborough [26]	2250	-

Table 2.2 - Inlet mass velocities of some TOTAL vacuum heaters[1]

Refinery	**Mass velocity kg/m^2.s**	**Feed rate ton/h**
Donges	1700	520
Feyzin	2690 (1620-2890)	230 (139-248)
Grandpuits	950 (760-1218)	260 (200-320)
Flandres	1237 (726-1560)	230 (135-290)
Provence	1105 (800-1380)	300 (210-350)
Leuna	1606	645
TRN	1730	322
Normandy D8	1255	117

[1] Values in parenthesis represent operation intervals due to feed rate change. The value outside the parentheses is the most common. Values were obtained from references [31] [32], calculations and FRNC-5 simulations.

Recommended minimum mass velocities used by refiners are all similar, but the values recommended by some engineers in scientific articles are much higher.

When looking at the mass velocities of vacuum heaters currently in operation at TOTAL we see two things, that there are very different values and that for a given heater, turndown rates can be much lower than the usual value.

With the exception of Grandpuits, all the other vacuum heaters from former Elf refineries: Donges, Leuna and especially Feyzin, have higher mass velocities than TOTAL heaters. The Feyzin and Donges vacuum heaters have some of the longest run lengths in the group according to a Benchmarking study [32], which may indicate the influence of high mass velocity in the coking rates. There are two other heaters with high mass velocities. For the TRN heater, run length information is not available. As for the Leuna vacuum heater it has a geometry that results in very high localized heat flux, causing coking problems in some parts of the tubes, this will be discussed in section 6.1 of this report.

2.1.2.FRNC-5 simulation study

The importance of mass velocity in limiting coke formation in fired heaters has been well established but it is not quantified. Why do refiners use 1250 kg/m^2.s? Why not use a higher mass velocity since it can improve run length, as the data from TOTAL heaters seem to suggest?

To have some idea of how the coking rate evolves with increasing mass velocity I used the fired heater simulation software FRNC-5. This software calculates a relative coking rate that can be used to compare different conditions. This relative coking rate is influenced by two parameters, film temperature and residence time. There are other things that influence the coking rate, such as wall sheer stress and charge quality, but they are not as influential as film temperature and residence time, which incorporate the changes in several heater design and operational parameters.

I simulated a heater with only one radiant tube per pass and maintained the same outlet temperature, changing only mass velocity. The simulations were repeated at different heat flux levels, since it also has an important influence on film temperatures and therefore on coking.

My results of FRNC-5 simulations with varying mass velocity are shown in figures 2.1 and 2.2. Figure 2.1 shows the results of simulations made with a 385 ºC outlet temperature and no vapor in the tubes. Figure 2.2 shows the results of simulations made with a 415 ºC outlet temperature and at 9% wt vapor at the outlet with at least 5% wt vapour at the inlet. The results show that when there is no vaporization in the tubes, mass velocity always decreases the coking rate but its effect is much less marked after 1500 kg/m^2.s. When vaporization occurs in the tubes, high mass velocities change the two-phase flow regime, which in turn increases film temperatures and therefore the coking rate.

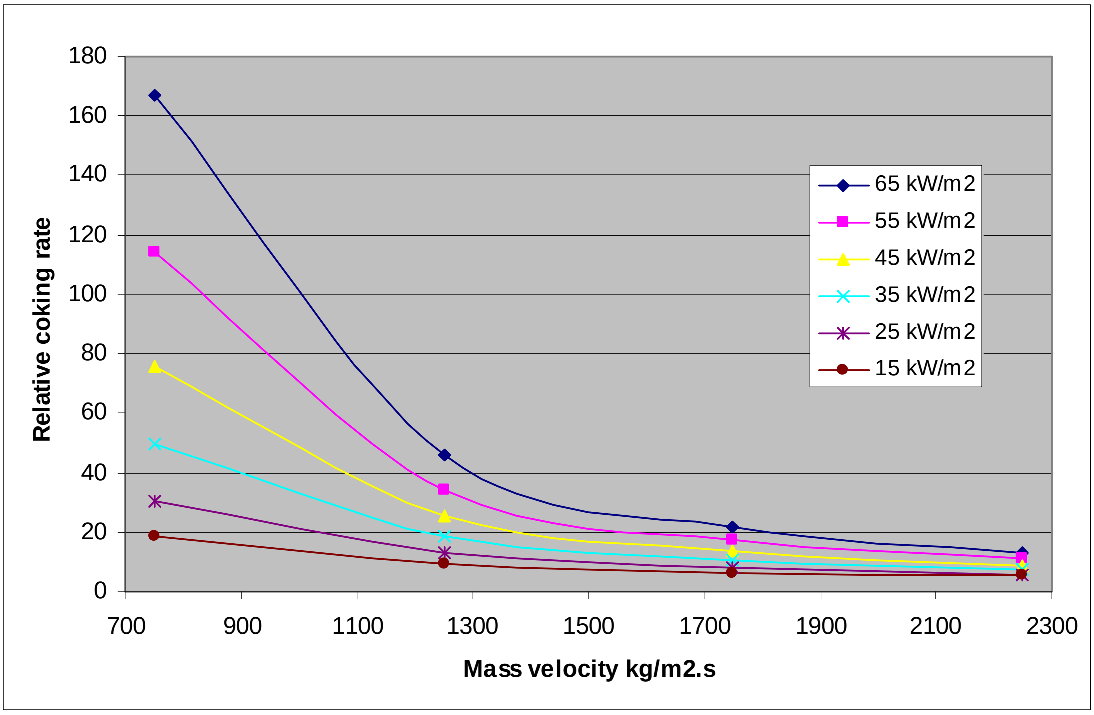


Figure 2.1 - Coking rate variation with mass velocity at different heat flux with no vaporization

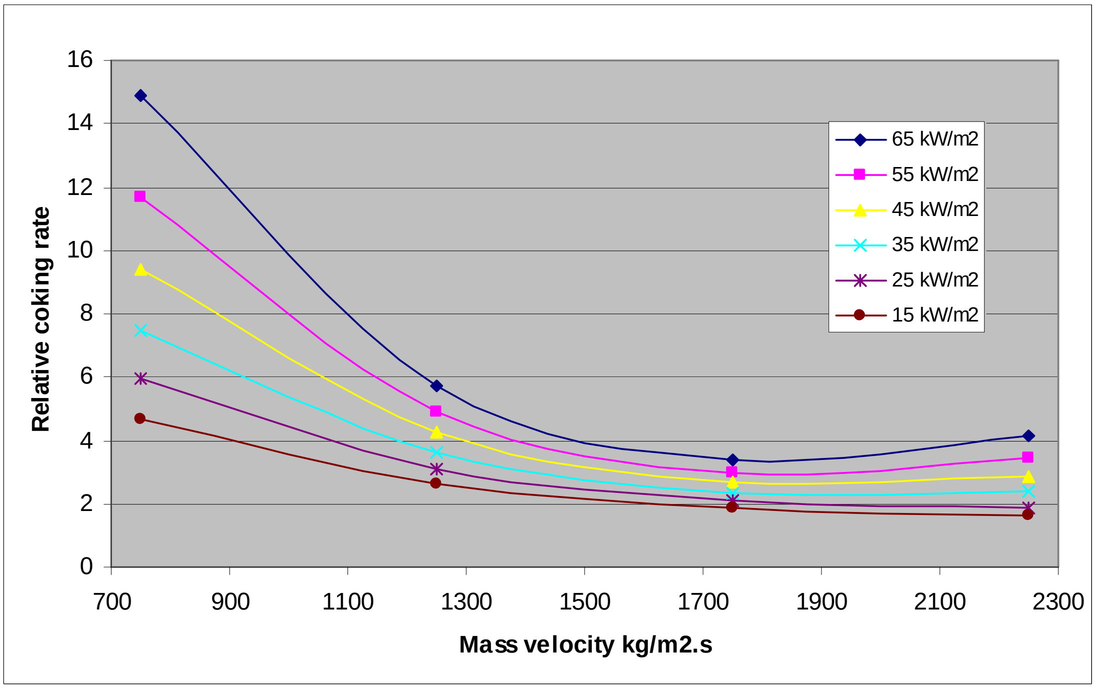


Figure 2.2 - Coking rate variation with mass velocity at different heat flux with 9% wt vapor

The outlet temperature was found to have no effect on the relative variation of the coking rate with mass velocity (results obtained at different outlet temperatures presented the same curve tendency), only on the absolute values of the coking rate. On the other hand, vaporization has a marked effect on the coking rate, lowering it considerably.

We can see that both curves show inflections between 1500 and 1750 $kg/m^2.s$. The best mass velocity value should be in this interval, since using mass velocities higher than 1750 $kg/m^2.s$ decreases the coking rate only slightly and it can even increase, if a two-phase flow regime change occurs. For mass velocities below 1500 $kg/m^2.s$ there is still a large coking reduction potential that isn't explored, as shown by the curves in figure 2.1.

The main problem with increasing mass velocity is additional pressure drop. Figure 2.3 shows the linear variation in inlet pressure with mass velocity in the Leuna VDU heater, obtained by FRNC-5 simulations. We can see that a 40% increase in mass velocity increases the pressure drop by 40%. This is why we can't use very high mass velocities and the 1500 $kg/m^2.s$ to 1750 $kg/m^2.s$ mass velocity interval is recommended. Because these mass velocities provide a significant coking rate reduction without too high an increase in pressure drop.

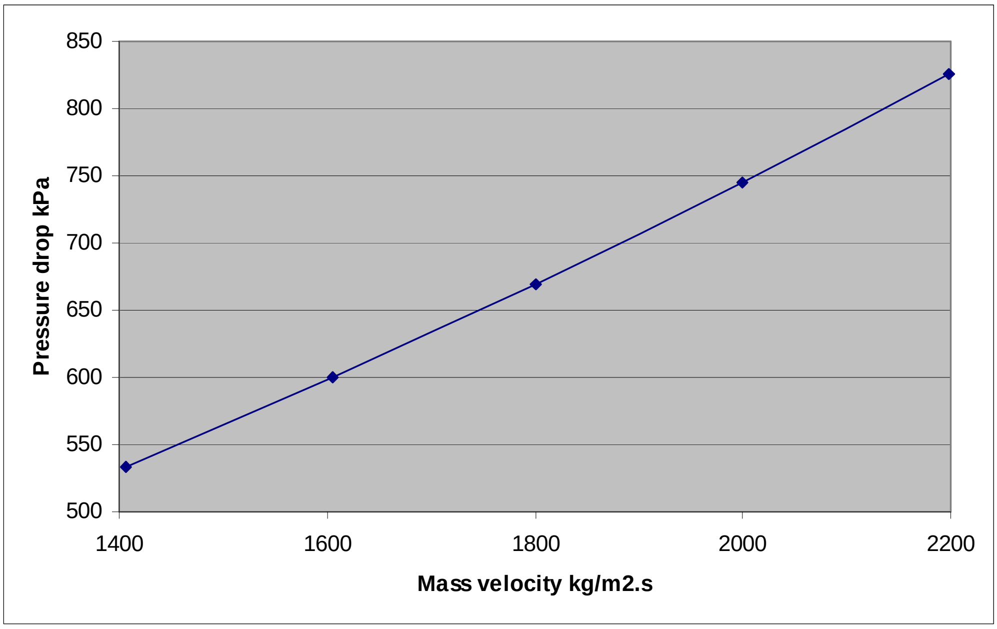


Figure 2.3 - Inlet pressure variation with mass velocity in the Leuna VDU heater

To recommend a single value lets analyze the interval. The coking rate reduction between 1500 $kg/m^2.s$ and 1750 $kg/m^2.s$ is much less significant than the one between 1250 $kg/m^2.s$ and 1500 $kg/m^2.s$ and pressure drop increase is linear. **Therefore I recommend using 1500 $kg/m^2.s$ as the design mass velocity for new vacuum distillation unit fired heaters**.

2.2. Film temperature

Film temperature is at least or even more important a design criterion than mass velocity because film temperatures are the highest fluid temperatures in a heater. Since the coking rate has an exponential dependence on temperature a small increase in temperature means a big increase coking.

The coking rate approximately doubles with every 20 ºC increase in temperature. By specifying a maximum film temperature we can make sure the coking rate will stay within acceptable values, provided a minimum fluid velocity and adequate vaporization are guaranteed to keep residence time low. That is why the three most important design parameters also include mass velocity and coil steam injection.

The difference between specifying a minimum mass velocity and a maximum film temperature is that, while it's easy to specify an inlet mass velocity for a heater, the film temperature changes throughout the heater's tubes. So it must be calculated for every tube in the heater while considering flow in the tubes and heat flux distribution in the firebox.

Balancing the heat flux is the best way to lower oil film temperature. This can only be achieved with careful and detailed sizing of the heater which usually collides with the refiner's desire to save in investment capital.

2.2.1.Collected values

Film temperature at the tube wall is calculated from the fluid bulk temperature using the heat flux and the heat transfer coefficient.

$$T_{film} = T_b + \frac{\phi}{h_{film}} \qquad (2.1)$$

It is worth noticing that the outlet bulk temperature is not the highest bulk temperature inside the heater. As it happens in the transfer line, in the heater pressure drop causes the oil to vaporize. This happens faster than the heater can provide heat to the oil, so an isenthalpic expansion occurs and the oil vaporizes by lowering its temperature. This is what happens in the transfer line and column flash zone inlet and it is for this reason that the heater's outlet temperature is always much higher that the column flash zone temperature. It also means that as the oil vaporizes towards the exit it will reduce its bulk temperature, so the point of highest bulk temperature in the heater is not the heater outlet but a point in the outlet tube or one or two tubes before.

The accepted and most widely used method to calculate the heat transfer coefficient is the one described in API Standard 530 [33]. The relevant equations are reproduced below.

Equation 2.2 is valid for liquid flow with Re>10 000 and equation 2.3 is valid for vapor flow with Re>15 000. In equation 2.3 the temperature is used in Kelvin. Laminar flow is rare in process heaters and the currently available correlations are not adequate for this type of service, so only correlations for turbulent flow are given.

$$h_l = 0.023\left(\frac{k_b}{D_i}\right)\mathrm{Re}^{0.8}\,\mathrm{Pr}^{0.33}\left(\frac{\mu_b}{\mu_w}\right)^{0.14} \qquad (2.2)$$

$$h_v = 0.021\left(\frac{k_b}{D_i}\right)\mathrm{Re}^{0.8}\,\mathrm{Pr}^{0.4}\left(\frac{T_b}{T_w}\right)^{0.5} \qquad (2.3)$$

These equations reproduce experimental data within ±20%, so this should be taken into consideration when calculating film temperatures. Also the Reynolds number should be calculated using the fluid's total mass velocity, not just the mass velocity for liquid or vapor. And the properties used to calculate the heat transfer coefficient should be for liquid or vapor depending on which coefficient is being calculated.

Re and Pr are the Reynolds and Prandtl numbers calculated with equations 2.4 and 2.5.

$$\mathrm{Re} = \frac{D_i Q_{mA}}{\mu_b} \qquad (2.4)$$

$$\mathrm{Pr} = \frac{Cp\mu_b}{k_b} \qquad (2.5)$$

For two-phase flow, equation 2.6 should be used to approximate the heat transfer coefficient. If dispersed or mist flow regimes occur the heat transfer coefficient should be calculated with the correlation for vapor phase of equation 2.3, rather than approximating it with equation 2.6.

$$h_{2p} = h_l w_l + h_v w_v \qquad (2.6)$$

This method is relatively accurate except when used to estimate heat transfer coefficients in annular flow. Since annular flow predominates in most vacuum heaters' tubes, it is important to take a conservative look at the heat transfer coefficients calculated for this regime. Before calculating the film temperature the heat transfer coefficient should be decreased by 20%, as that is the maximum deviation in single-phase flow calculations and we lack a better error estimate for two-phase flow calculations.

A quick and easy way of estimating film temperature, suggested by PCS, is adding 15 to 45 ºC to the bulk temperature. But this is to be used only for estimates, not for heater design.

At temperatures below 420 ºC coking will be very slow, so the problem is film temperatures above this value. Shell recommends keeping film temperature below 460 ºC to allow run lengths of up to 4 years. Heurtey Petrochem permits calculated film temperatures of up to 480 ºC, but their heat transfer coefficient calculation method is more conservative than the one recommended by API. As a note, the new RN VDU has been designed with a maximum film temperature of 454 ºC.

Using these recommended values and taking into consideration that the error of the two-phase heat transfer coefficients calculated by the API 530 [33] method is not known, **I recommend using 460 ºC as the maximum film temperature allowable in vacuum heaters**. But since we know that the single-phase heat transfer coefficient correlations have maximum errors of ±20%, it would perhaps be wise to allow for this error in film temperature calculations.

This rule does not apply to heaters' outlet tubes with mist two-phase flow regime. Due to the fluid contacting the walls being essentially vapor, film temperatures are usually much higher than 460 ºC. This won't be a problem if the fluid velocity is high, as described in section 2.7 of this work.

2.2.2.FRNC-5 simulation study

FRNC-5 simulations permitted an analysis of the behaviour of the coking rate with changing temperature. Film temperatures are mostly responsible for coking so these are the temperatures used in this study. Figure 2.4 shows the coking rate exponential dependence on temperature at two different mass velocities.

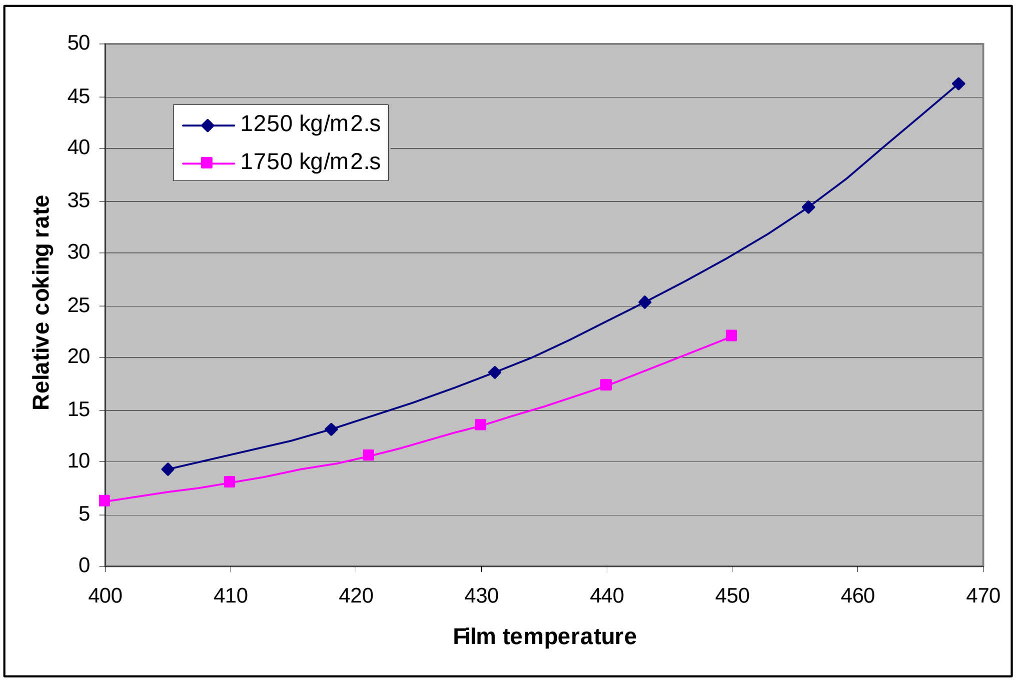


Figure 2.4 - Coking rate variation with film temperature

2.3. Coil steam injection

2.3.1.Injection point

Coil steam is used to promote charge vaporization. It lowers the hydrocarbons' partial pressure inducing vaporization at lower temperatures, which limits the fluid bulk temperatures and therefore lowers the film temperature. The added volume raises the fluid velocity in the tubes, this leads to higher heat transfer coefficients and wall shear stress and lower residence times, if a two-phase flow regime change does not occur. All this helps to lower coking rates.

Water is in vapor phase at the conditions present in fired heaters and it will lower the hydrocarbons' partial pressure promoting charge vaporization. When a fluid starts vaporizing it will vaporize at the tube wall, where fluid temperature is highest. But vaporization means the fluid has reached a saturated state and that additional heat received will be transformed into latent heat (vaporization of the fluid) instead of sensible heat (fluid temperature rise). So the film temperature will be limited to the fluid's saturation temperature at the pressure in the tube.

The coil steam injection point should be chosen before high coking temperatures. A good rule is to **inject the coil steam before the first tube where fluid bulk temperature reaches the heater's outlet temperature**, which is around 420 ºC for new heaters.

It should be injected at the inlet of the tube before the one where fluid bulk temperature reaches the heater's outlet. This is because the injection of steam in the coils creates an unstable two-phase annular flow that takes two to three tubes to dissipate. Injecting too much steam will produce an annular flow that will leave the liquid creeping along the tube walls, while vapor flows at high velocity in the center of the tube.

An additional coil steam injection can be made downstream to avoid high film temperatures. This is preferable to injecting all the steam at once because of the additional pressure drop it causes. Figure 2.5 shows the influence of coil steam in pressure drop when oil vaporization occurs.

2.3.2.Coil steam rate

The injected coil steam should allow enough vaporization to keep the fluid bulk temperature below the heaters outlet temperature, while maintaining film temperatures low. A normal initial vaporization is 3% to 4% mass fraction of the charge.

Usual steam feed rates are between 0.2% and 0.5% wt of the oil feed rate. More than 1% wt of the oil feed rate is unusual, although it may sometimes be necessary. Very high steam feed rates cause elevated pressure drop and may also develop annular flow, causing the vapor to flow very fast in the center of the tube while leaving the oil creeping along the walls with high residence time. The added pressure drop created by the steam injection increases the boiling temperature of the fluid, in effect increasing film temperatures.

2.3.3.FRNC-5 simulation study

AS this is a complicated matter, I thought best to study the influence of coil steam through FRNC-5 simulations. Simulations of a 14 meters tube with vaporization were used to study the influence of coil steam on pressure drop. The results are presented in figure 2.5, where we can see that the pressure drop has an almost linear variation with coil steam. The relation is not completely linear only because of vaporization occurring inside the tube.

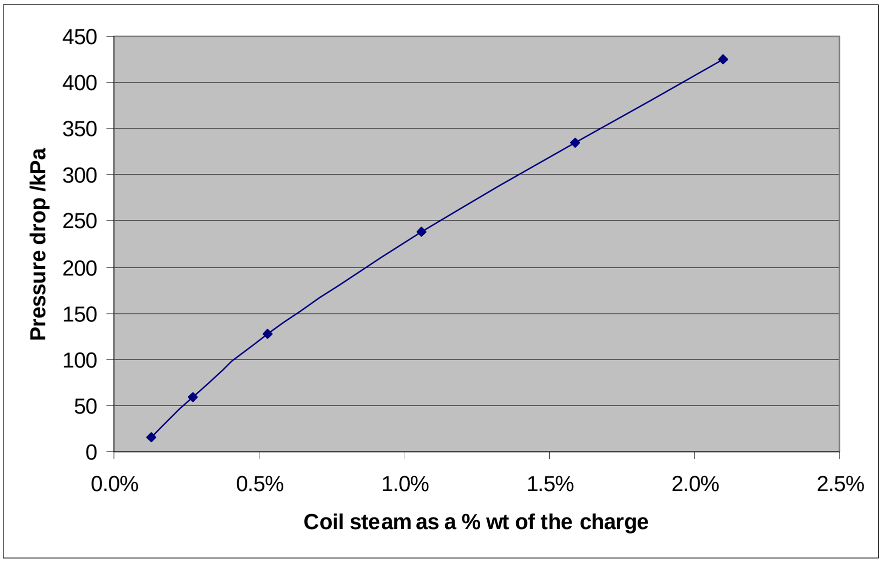


Figure 2.5 - Pressure drop as a function of coil steam injection at 1750 $kg/m^2.s$ mass velocity

More than one coil steam injection can be in different points of the tube coil. When a lot of coil steam is used, typically more than 0.5% wt, this is preferable to injecting all the steam at once because of the additional pressure drop it causes.

A more important study was the influence of coil steam on the coking rate. Operators sometimes increase the coil steam rate to try to compensate for low charge rate or rapidly increasing skin temperatures. The same model used above was used and the results can be seen in figure 2.6, which shows the coking rate reduction as more steam is injected, with a heavy residue being processed at 1750 $kg/m^2.s$ mass velocity. It is clear that in these pressure and temperature conditions and for this particular residue, injecting coil steam at more than 0.5 %wt of the oil feed rate does not improve conditions in the tube. The coking rate will remain the same with the problem of added pressure drop.

What we cannot see in the graph is that injecting coil steam at more than 0.5% wt of the oil feed rate, changes the two-phase flow regime to mist, which results in dry walls and high film temperatures.

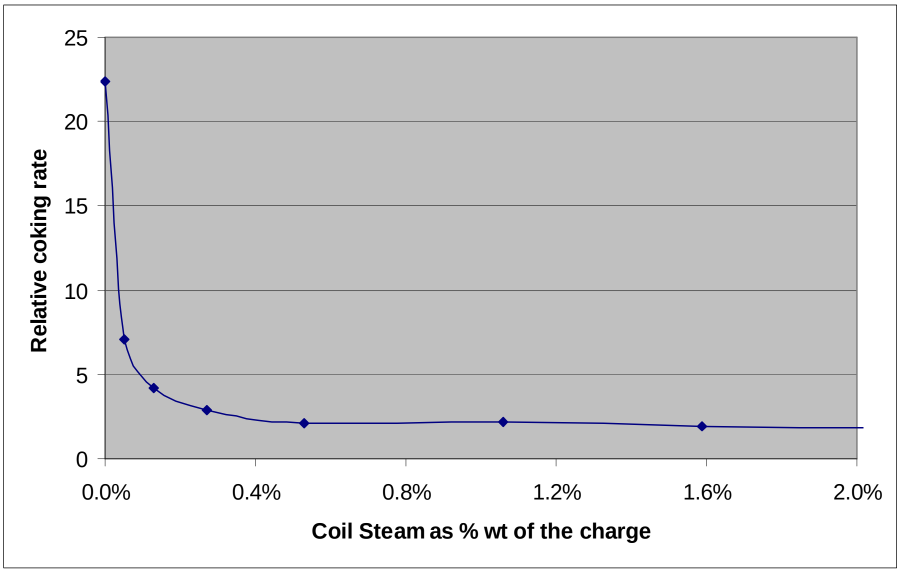


Figure 2.6 - Coking rate vs coil steam rate with a heavy residue at 1750 $kg/m^2.s$ mass velocity

The conclusion is, that the main film temperature lowering effect is not produced by the fluid speed created by the steam. Therefore, **if injecting more steam does not promote additional vaporization it should not be used**. There is another downside, more injected steam means more has to be condensed and more motive steam has to be used in the vacuum column ejectors, limiting pressure in the column and in turn increasing pressure in the heater.

Of course each residue has different vaporization curves and each tube has its own pressure, temperature and vaporization conditions. Therefore the **adequate coil steam rate has to be evaluated on a case by case basis**. It should be studied for every heater and for each residue to be processed in the conditions the heater will operate in. But it is important to remember that there is always a limit to how much good injecting more coil steam can do.

2.4. Heat flux

We have seen the importance of restricting fluid film temperature and the way to do it is by balancing heat flux [34]. **High localized or peak heat flux causes elevated tube metal temperatures even when no coke is present** [35]. This often raises the oil film temperature above the oil's thermal stability and causes coke to form. In figures 2.1 and 2.2 of section 2.1 we saw the effect of heat flux on the coking rate, especially in low mass velocity conditions.

Heat flux isn't really the problem here, bad convection inside the tubes and coke thermal resistance are because they limit heat transfer. Film heat transfer coefficients are the limiting factor in fired heater heat transfer. They are very dependent on flow regime, increasing with Reynolds number. They are also extremely dependent on two-phase flow patterns, as can be seen in section 2.12. According to UOP, film heat transfer coefficients are usually between 1700 W/m^2.K and 2850 W/m^2.K but in some situations they can be lower than 1000 W/m^2.K.

Several factors including the number of burners, burner layout, burner flame length, burner to tube spacing, firebox height to width ratio and non-ideal flue gas patterns influence heat flux distribution. Since heat transfer is essentially made by radiation in the radiant zone, the proximity to the flame determines the actual heat flux received by any part of the tube. Also, parts of the tubes that don't have direct view of the flame, receive radiation reradiated by the refractory walls. Due to this difference in radiation source and direction, the front face of the tube receives as much as 6 times more heat than the back. The circumferential heat flux distribution for single and double-fired tubes is shown in figure 2.7.

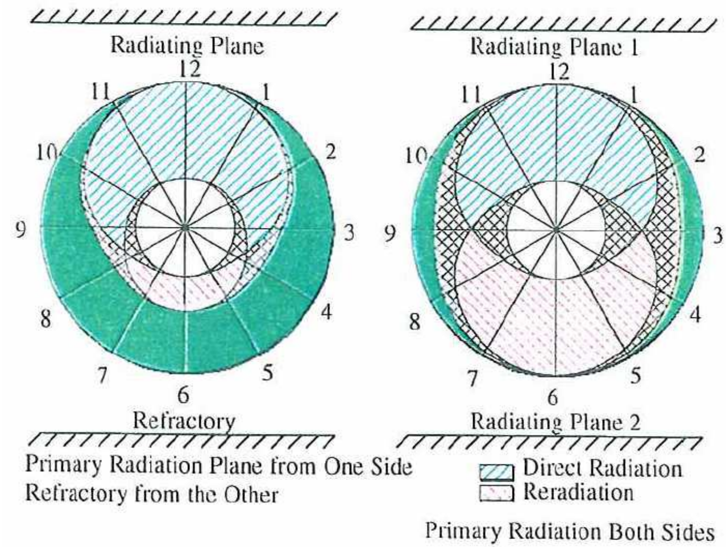


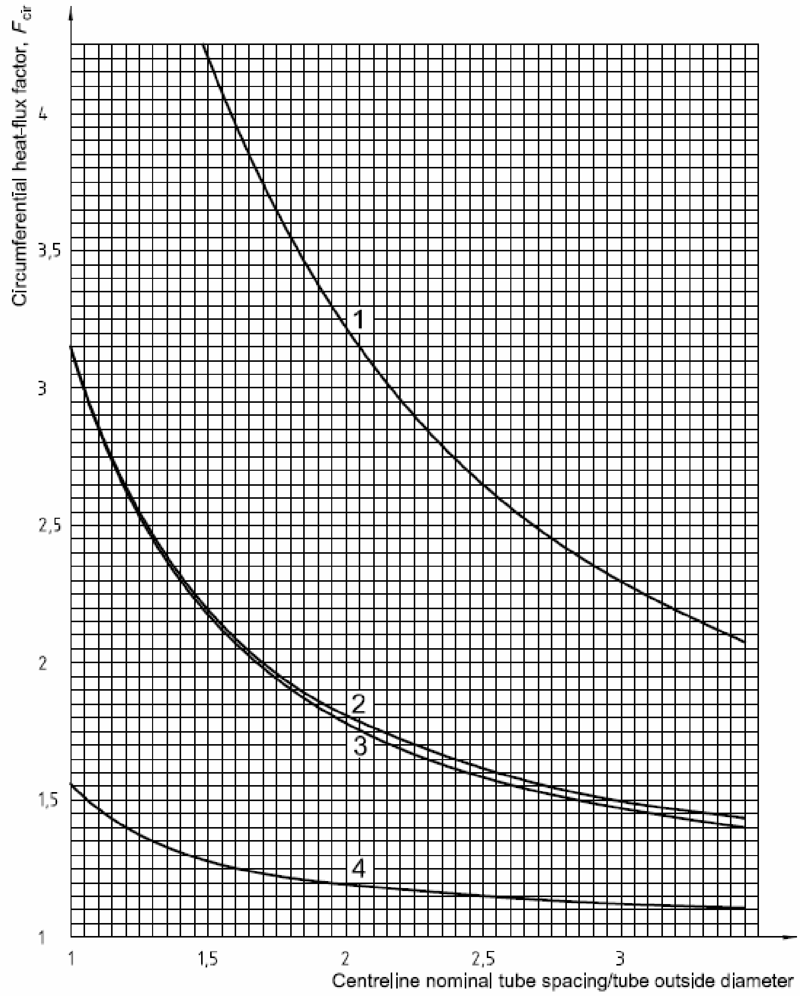
Figure 2.7 - Heat flux distribution around tubes [35]

Peak heat flux can be estimated from the heater's average radiant heat flux because for each type of heater configuration the approximate heat flux distribution is known. Equation 2.7 allows us to calculate the peak heat flux using factors that account for circumferential and longitudinal heat flux variations.

$$\phi_{peak} = \phi_{ave,rad} F_{cir} F_L + \phi_{conv} \qquad (2.7)$$

Local heat flux densities vary considerably throughout a heater because of non uniformities around and along each tube. Circumferential variations result from differences in radiant heat flux density produced by shading from other tubes or from the placement of tubes next to a wall. Conduction around and along the tube walls and convection heat transfer from flue gas, tend to reduce the circumferential variations in heat flux density. Longitudinal variations result from: proximity to burners, flame heat flux profiles and the bulk fluid temperatures. In addition to variations in the radiant section, tubes in the shock section of a heater can have high convective heat flux density.

The circumferential variation factor, F_{cir}, is given as a function of tube spacing and coil geometry in Figure 2.8. The factor given by this chart is the ratio of maximum local heat flux density at the fully exposed face of a tube, to average heat flux density around the tube. This chart was developed from considerations of radiant heat transfer only. Influences such as conduction around and along the tube and flue gas convection, act to reduce this factor. Since these influences are not included in this calculation, the calculated value will be somewhat higher than the actual maximum heat flux density.



Key

1 Curve 1 = double row against wall, triangular spacing

2 Curve 2 = double row with equal radiation from both sides and two diameters between rows, equilateral spacing

3 Curve 3 = single row against wall

4 Curve 4 = single row with equal radiation from both sides

Figure 2.8 - Ratio of maximum local to average heat flux around a tube [33]

Also, the curves in figure 2.8 are valid when a tube center to refractory wall spacing of 1.5 times the nominal tube diameter is used. This is the usual tube center to refractory wall spacing but if a different spacing is used these curves will shift a little, depending on increasing or decreasing tube to wall spacing.

The circumferential heat flux variation factor for the most usual tube arrangements is 1.9 for single-fired tubes, spaced two tube diameters from tube center to tube center and 1.2 for double-fired tubes with the same spacing.

The longitudinal variation factor, F_L, is not easy to quantify. Values between 1.2 and 1.3 are the most common, although it can vary between 1.0 and 1.5. For gas firing 1.2 is usually used as is 1.3 for oil firing, in heaters with an L/D around 2. In a firebox that has a very uniform distribution of heat flux density, a value of 1.0 can be appropriate. Values greater than 1,5 can be appropriate in a firebox that has an extremely uneven distribution of heat flux density (for example, a tall, narrow firebox).

Figure 2.9 shows the heat flux imbalances due to the form and fuel nature of the flame, which greatly influences the longitudinal variation factor. The longitudinal peak heat flux is generally located at 1.5 m when burning oil and 3.5 m when burning gas, but it will be higher if low NOx burners are used.

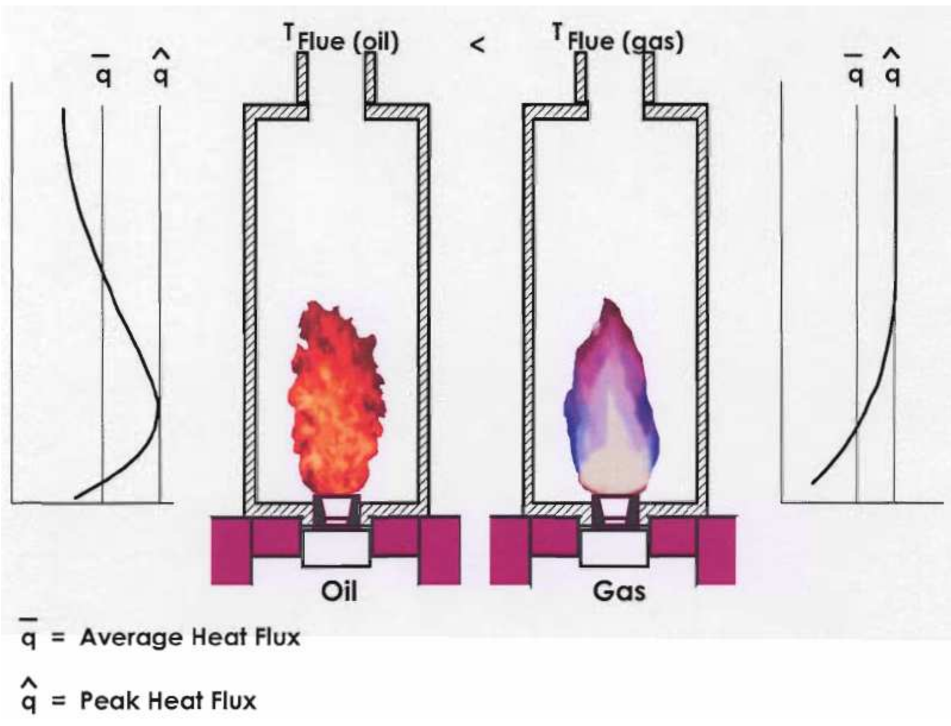


Figure 2.9 - Burner flame heat flux profiles [9]

The peak heat flux is the highest heat flux expected in the heater's tubes and can be as high as 2.5 times the average heat flux due to heater geometry. If flame impingement occurs it can be even higher. The peak to average heat flux ratio is around 1.8 to 2 for most heaters.

These methods for calculating the peak heat flux have more than 40 years, but even before them, heaters were designed for a certain average radiant heat flux, usually based on experience. Specifying the average heat flux has traditionally been the way to design fired heaters because by limiting the peak heat flux in the heater we can limit the peak film temperature.

The problem is film temperature also depends on bulk temperature and so, if a high heat flux affects a tube near the heater's outlet a coking problem will develop. Film temperature should be the design consideration to meet, not heat flux. This is not to say the average heat flux concept is useless. When designing a fired heater we must have a starting point. We can start by designing a heater with a certain average heat flux and then, using the variation factors or CFD software (for example Xfh or FLUENT) we can calculate film temperatures in every tube in the heater to verify if they respect the guideline.

There are lots of opinions on what the maximum average heat flux should be for vacuum heaters. TOTAL sets a maximum of 26 kW/m^2 for single-fired tubes and 38 kW/m^2 for double-fired tubes in vacuum heaters. UOP recommends the average heat flux for single-fired tubes in vacuum heaters to be 25 kW/m^2. Shell gives 50-60 kW/m^2 as the acceptable peak heat flux for vacuum heaters [9], if we use a peak to average heat flux ratio of two, the average heat flux ratio will be 25-30 kW/m^2.

Heurtey Petrochem designs vacuum heaters with 22 kW/m^2 to 28.5 kW/m^2 average heat flux for single-fired tubes in lubricant bases production and 28.5 kW/m^2 to 35 kW/m^2 average heat flux for single-fired tubes in fuel production service.

Several other values can be found in the literature for single-fired tubes in vacuum heaters, most of them between 25 kW/m^2 and 35 kW/m^2. Most engineers consider 25 kW/m^2 to be a moderate heat flux and 35 kW/m^2 a rather high heat flux, but still acceptable if the heater has a mass velocity higher than 2000 kg/m^{2}s and uses coil steam.

Double-fired tubes have much lower peak to average heat flux ratios which allows us to use a higher average heat flux. Most people agree with using 1.5 times the average heat flux for single-fired tubes.

If we consider all these opinions **the current TOTAL specification of 26 kW/m^2 for single-fired tubes and 38 kW/m^2 for double-fired tubes seems reasonable** [36]. When designing a heater this can be a good starting point. After calculating film temperatures, if they are too high or too low, the heater design and the average heat flux can be adjusted accordingly.

2.5. Type of heater

Heaters are classified according to three characteristics, structural configuration, radiant tube coil configuration or shape and burner arrangement.

Typical structural configurations are cylindrical, box, cabin and multi-cell box. **The typical radiant tube coil configurations in vacuum heaters are vertical and horizontal** but helical and arbor exist in other types of heaters. **Burner arrangements include: upfired**, (burners on the heater's floor), **downfired** (burners on the heater's arch) and **wallfired** (burners on the heater's walls). The wallfired arrangement can be further classified as endwall fired, sidewall fired and sidewall fired multi-level. Typical heater types are shown in figure 2.10.

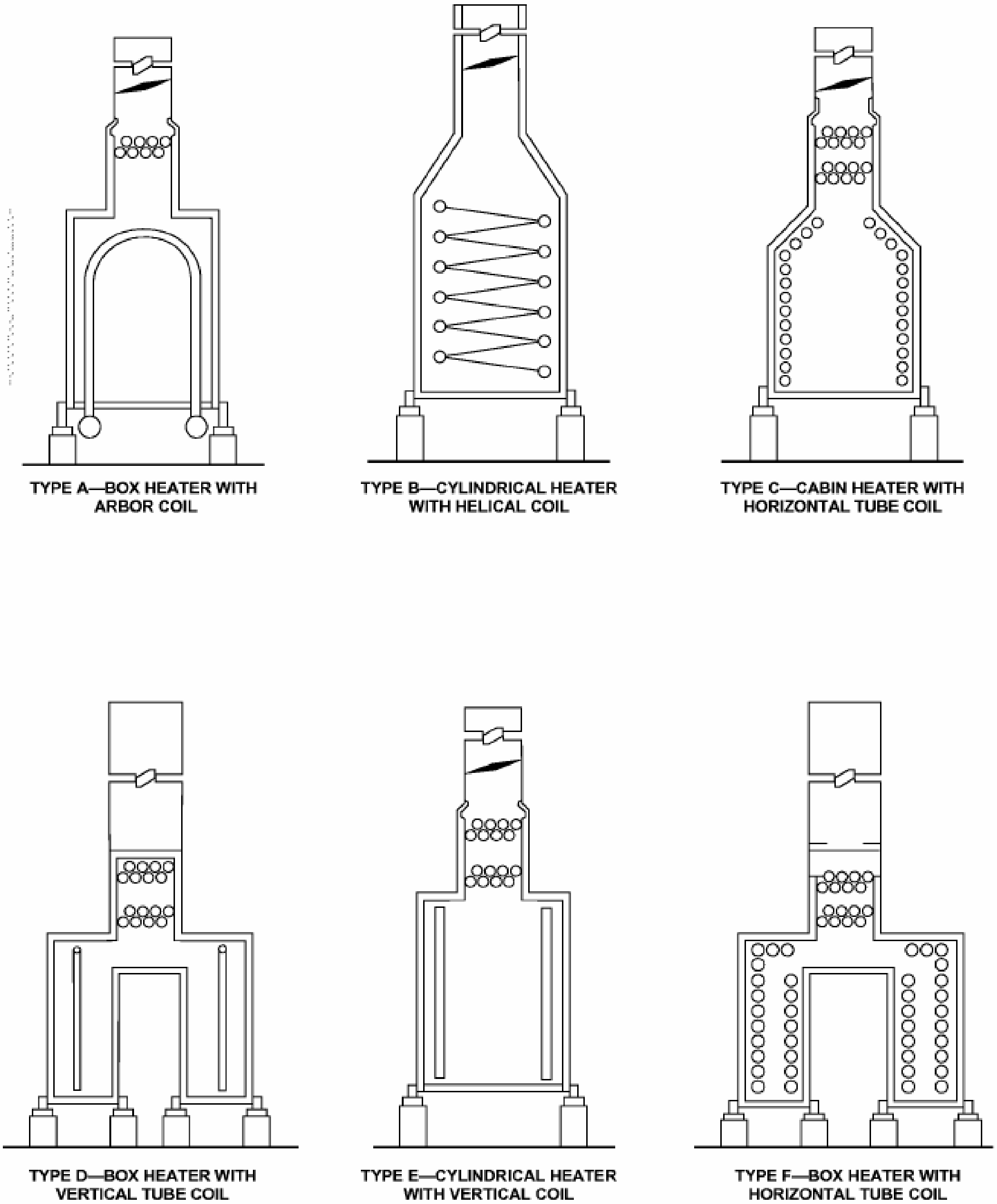


Figure 2.10 - Typical heater types [1]

The type of heater to choose depends essentially on heater's feed rate and firing duty, which depends on inlet and outlet temperatures and charge vaporization. This along with maximum allowable pressure drop will define the number of passes which is determinant in choosing the heater configuration.

For heaters with duties up to 50 MW cylindrical heaters are cheaper to built, they also occupy a smaller land plot. Cylindrical heaters have vertical tubes and allow the use of bigger burners because the radial heat flux distribution in these heaters is more uniform than in box heaters. This also allows more tube passes since all tubes receive the same heat flux. The disadvantages of these heaters are higher probability of flame impingement due to the use of bigger burners and high vertical heat flux imbalance.

For heaters with duties above 50 MW a cabin or box heater is preferred because it is cheaper than a 2 cell cylindrical heater. These heaters can have horizontal or vertical tube coil configurations and can also have single and double-fired tubes.

Cabin and box heaters are usually built with 1 or 2 passes per cell with vertical tubes and 4 passes with horizontal tubes. If more passes are needed more cells can be used.

The height width/diameter ratio should be between 2 and 2.5 in cabin/box heaters and between 2.5 and 2.8 in cylindrical heaters to avoid high vertical heat flux imbalances and flame impingement [28] [36] [37]. Adequate burner to tube clearance shall be provided as described in section 2.6.3.

2.6. Tube coil arrangement

Despite of the chosen coil arrangement, the most important thing to consider is having the same heat flux to all passes, a pass being a single circuit without ramifications from inlet to outlet. **Passes must be balanced in order to receive the same amount of heat**. Tubes can be horizontal or vertical. Horizontal tubes with downward flow are the most common arrangement in service. New heaters usually use vertical tube arrangements because of its lower cost.

It is easier to balance vertical tube coils since all tubes receive approximately the same heat flux, especially in cylindrical heaters. Because of this and of the need for less tube supports, the vertical tube arrangement is cheaper than the horizontal. But vertical coils have a problem with vertical heat flux distortion that can place a high peak heat flux in all the tubes. As all tubes receive the same heat flux, the tubes with higher fluid bulk temperature will suffer from elevated film temperatures in the area of the tubes subject to high heat flux.

Another downside to using vertical coils is the difference in hydrostatic pressure between the top and bottom of each tube. This may lead to flow regime changes inside the same tube and from one tube to the next. The tube bends and the tube sections after each bend are especially prone to such behaviour. This may cause high film temperatures leading to coke formation, a 10 to 15 ºC temperature difference between adjacent tubes with ascending and descending flow is not uncommon and is sometimes enough to cause coking problems. Heaters with long vertical tubes (60 ft) have coking problems in the top part of the return tubes (tubes with down-flow) due to this. Some companies limit vertical tubes to 15 m (50 ft) to avoid this problem and high longitudinal heat flux variation factors.

Horizontal coils allow for the vertical adjustment of tubes with cold oil to high heat flux areas. The cold oil from the convection section can be fed to the tubes that receive the highest heat flux. Tubes near the heater's outlet have high vaporization fractions with high oil residence times and low mass velocity, which is a situation very favourable to high coking rates. They should be placed in a zone with low heat flux, such as near the arch or near the floor. This prevents the development of high film temperatures in these tubes.

Another concern with horizontal coils is pass stacking when 4 or more passes are put into the same heater cell. If passes are built with all tubes from the same pass stacked on each other one pass will be placed in the zone of vertical peak heat flux and the other(s) in zones with low heat flux. This will result in pass outlet temperature differences and a very hot pass with severe coking problems.

To avoid this, when opting for horizontal passes, the passes should not be stacked but instead the n^{th} tubes of each pass should be stacked. Only after all the first tubes of all the passes are stacked, should the second tubes be stacked. This can be altered to placing 2 or 3 tubes from each pass together to avoid high heater costs but even then this type of coil is much more expensive than stacking the whole pass and certainly much more expensive than a vertical coil, due to the added cost of tube crossovers versus the cost of simple return bends.

A well distributed horizontal coil with balanced passes is the best tube coil arrangement to avoid coking but it is also the most expensive. Because of the cost of a well balanced horizontal coil and the problems it may have if it is not properly balanced, almost all new heaters are built with vertical coil arrangements.

In vertical coil arrangement there is no general flow direction, tubes have alternately up-flow and down-flow. However, with a horizontal coil arrangement there is a difference. Because of the two-phase flow and high vapor volume in the exit tubes, it is advantageous to place them at the top of the heater to avoid height differences between the heater outlet and the column entrance. Vertical two-phase flow may lead to liquid slip and consequent higher pressure drop.

More information on this subject can be found on TOTAL Specification GS RM THE 001 [36], International Standard ISO 13705, API Standard 560 [1] and [38].

2.6.1.Number of passes and tube size

The most important variables when sizing heater tubes are mass velocity and pressure drop. Attention must also be given to flow regime when in two-phase flow. The number of passes and the first tube size are chosen according to these criteria.

Increasing the number of passes decreases coil pressure drop but it will result in higher heater costs. It can also cause more heat flux imbalances if proper care is not taken in heater design. But more passes mean smaller diameters and more tubes. This result in a smaller global volume, which can lower the residence time if we keep the fluid speed, or it can lower the pressure drop if we keep the oil residence time.

The usual radiant section inlet nominal tube size is between 3.5” and 6”, these values should be used with mass velocity to determine the number of passes that result in an appropriate heater pressure drop.

In vacuum heaters an important part of the charge is vaporized in the tubes, this changes the fluid specific volume and therefore increases the fluid velocity. In order to avoid prohibitive pressure drop and having the fluid travelling at erosion speeds, the tube size must be changed in the tubes where significant vaporization occurs so that fluid velocity is maintained within acceptable values.

Tube size is usually increased incrementally in the last 2 or 4 tubes before the outlet to avoid these problems. If high vaporization starts earlier it may be necessary to increase the tube size in previous tubes and use more tube sizes. The outlet tube size is generally limited to 10”-12” because larger tubes present problems of supporting and handling inside the heater [37].

Pressure drop in the heater should not be too high to avoid high vaporization temperatures and costs with pumping and the pre-heat train exchangers. UOP recommends a 5 bar pressure drop through the heater [30] but this is for an inlet mass velocity of 1220 $kg/m^2.s$. As this work recommends an inlet mass velocity of 1500 $kg/m^2.s$, **the pressure drop through the heater should be around 6 bar**. But it can be higher depending on charge vaporization and the use of coil steam.

UOP recommends 20 ft as the minimum vertical tube length in order to have a reasonable Bridgewall temperature at the firebox exit [30]. The maximum tube lengths are 60 ft for vertical tubes because of problems supporting longer tubes. But as we saw, other problems suggest limiting the tubes to 50 ft. Horizontal tubes are usually limited to 100 ft. [39]

Tube equivalent lengths with approximately the same pressure drop should be arranged for all tube passes, so that it does not result in a difference of flow between passes. TOTAL specification GS RM THE 001 [36] states that the difference in pass arrangements should not result in a difference in flow greater than 5% of the rated flow value per pass.

2.6.2.Tube spacing

Tube spacing in fired heaters affects the circumferential peak to average heat flux ratio and the tube bank effectiveness, as can be seen in figures 2.8 (in section 2.4) and 2.11.

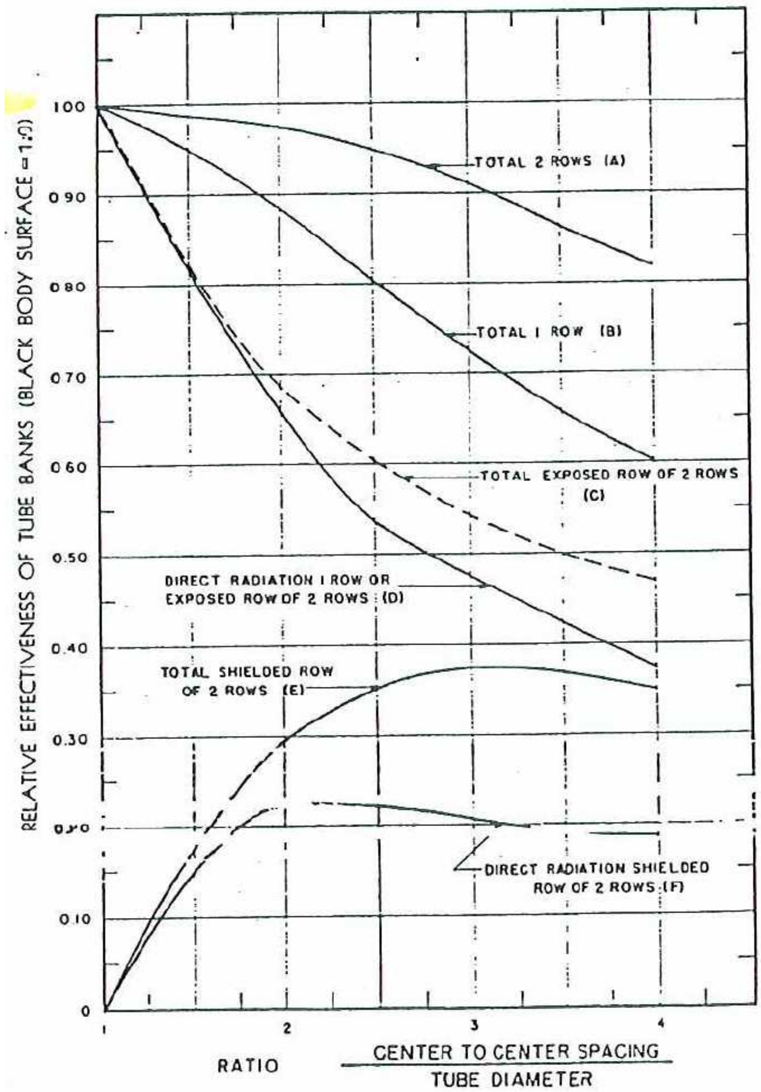


Figure 2.11 - Relative effectiveness of tube banks [10]

Changing tube spacing, number of rows and using single and double-fired tubes is an effective way of having different tube banks with different average heat flux in the same heater.

Using different tube spacing in the same heater is not very common nowadays, except for the last 2 or 3 tubes near the outlet, due to increased costs. Usual center to center tube spacing in most of the radiant section is 2 tube diameters, but 3 or 5 tube diameter spacing return bend sizes are also available in the market. Increasing the spacing lowers the peak to average heat flux ratio and it is used in tubes with large tube size. In the last 2 or 3 tubes near the outlet it is common to have a 3 tube diameters center to center tube spacing to avoid hot spots.

Increasing the usual 2 tube diameters spacing to 3 in all the tubes has a 50% higher return bend cost and also requires more refractory, amounting to a 25% increase in radiant section cost which represents 10% more in heater cost.

The tube center to refractory wall spacing used for single-fired tubes is between 1.5 and 2 tube diameters. Spacing between 1.7 and 1.8 tube diameters are the most common.

2.6.3. Burner to tube clearance

Special care should be taken with burners. Flame impingement can result in tube failure and high film temperatures from high peak heat fluxes, which are important causes of high coking rates. Burners should be in sufficient number and distributed evenly inside the heater in order to avoid high heat flux imbalance between sections of the heater's walls.

The various fuels used in fired heaters will exhibit different flame characteristics with, typically, heavy oils producing a longer flame than gas fuels. The flame length and diameter and the steadiness of the flame envelope will depend on: type of burner, firebox draft, excess air at the burner, flame interaction/merging and the amount of air leakage into the firebox.

In order to avoid impingement of the flame on heater tubes or refractory the industry has developed minimum standards for burner center line to tube center line clearance. These clearances are shown in figure A.1 in the annex section. Specifying a minimum clearance is necessary, because to be competitive a furnace vendor will normally base a price quotation on the smallest firebox suitable for the service. This is fine if the heater will always be fired with the burners properly cleaned and adjusted and with heat liberation not exceeding the maximum specified at any time. Because of their complexity, process heaters will not be firing at optimum conditions all the time.

Companies with extensive heater operating experience recognize that the API minimum clearances offer inadequate protection against common events such as burner fouling, operating upsets, rapid fuel and process changes, etc. Clearances in parentheses in column B of figure A.4 in the annex section reflect the practices of these companies. Figures A.2 and A.3 in the annex section show clearances recommended by UOP and engineers with extensive experience in heater design.

2.6.4.Tube metallurgy

Tube metallurgy has to be chosen according to the maximum skin temperatures expected in each tube at the end of each run length. Heaters are designed to have a certain run length at the expected coking rate at design conditions, so we can calculate an approximate value for the maximum expected coke thickness and from there the maximum skin temperature. If the heater fires fuel oil, a similar allowance must be given for scale deposits on the tubes exterior.

To the maximum calculated skin temperature, 30 ℃ s hould be added as recommended by TOTAL in its heater general specifications. Then a material, usually a chromium-molybdenum steel with 5% to 9% Cr, is chosen according to its limiting metal temperature. It is also necessary to calculate tube thickness according to tube elastic limits, creep-rupture limits and to the expected design life (TOTAL demands using at least 100 000 hours for this calculation).

More information on this subject can be found on International Standard ISO 13704 and API Recommended Practice 530 [33], which describes the complete procedure for heater tube thickness calculations.

2.7. Outlet tube conditions

Heater outlet temperature is adjusted according to desired VGO cut-point in the vacuum column. This will of course depend on column efficiency and transfer line pressure drop. It can be as low as 380 ºC or as high as 450 ºC but these are extreme cases. **Most existing vacuum heaters operate between 400 ºC and 430 ºC outlet temperature and recently designed ones between 420 and 430 ºC**. The outlet temperature depends on the type of charge to heat, temperatures higher than 420 ℃ are associated to particular crude oils and the injection of stabilization addictives is needed to slow coking.

PCS recommends a heater outlet pressure below 40 kPa [5] but this will also depend on the column flash zone pressure and transfer line pressure drop. This is not a reliable design criterion because heater pressure drop depends on charge vaporization. A very low heater outlet pressure will cause massive vaporization in the outlet tube, increasing the pressure drop in the heater. The design criterion to use should be heater pressure drop, if it is too high, increasing the heater outlet pressure a little by adjusting transfer line pressure drop may actually help. As an example, heater outlet pressures of 50 or 60 kPa are not uncommon in fact they may be more common than heater outlet pressures below 40 kPa. Heater outlet pressures are not usually monitored so they are mainly available from heater design calculations in equipment data sheets.

Has it was described in section 2.6.1 the last tubes in the heater have increased tube sizes due to high vaporization, this increases fluid velocity which is highest in the outlet tube. In the heater outlet, fluid velocity should be as high as possible to avoid coking but without exceeding the critical velocity. This would lead to high pressure drop due to choking. However, the outlet tube size is generally limited to 10”-12” because larger tubes present problems of supporting and handling inside the heater [37]. If necessary, more passes have to be used or transfer line pressure drop increased.

The high vapor fraction means mist two-phase flow regime is always present in the outlet tube of vacuum heaters. This regime consists of a gas phase with dispersed droplets of liquid. The absence of a liquid film contacting the walls leads to very high inside tube wall temperatures. When liquid droplets hit the wall they coke due to the high temperature, so coking rates in these tubes should be very high. Nevertheless, experience tells us coking problems can be avoided in these tubes if we keep high fluid velocity. The problem is erosion may occur if fluid velocity is too high.

In the outlet tube the fluid velocity should be 80% of the sonic velocity of the gas phase, which is sufficient to limit coking. Fluid velocities of 85% of the gas sonic velocity or above should not be used because they cause erosion problems, at least in most heater designs.

In transfer line design, fluid velocity should be limited to around 50% of the gas sonic velocity. This can also be used for the heater outlet tube but it is considered too conservative. Another criteria is the transfer line temperature drop, if it is higher than 20 ºC it is too high and the transfer line should increase in size. If necessary more than one transfer line can be built to avoid tube diameters above 52”, due to cost and stratification problems.

Critical velocity can be estimated using equation 2.8 where R can be assumed to be 0.333 for most refinery hydrocarbon streams. The critical velocity values calculated with equation 2.8 are not rigorous but they usually give satisfactory results. Another way to estimate the critical velocity is assuming it to be between 50% and 60% of the gas sonic velocity.

$$v_c = \frac{\alpha.\left[x + (1-x).\frac{\rho_g}{\rho_l}\right]}{\sqrt{x + (1-x).R^2.\left(\frac{\rho_g}{\rho_l}\right)^2}} \qquad (2.8)$$

In these tubes we need to worry only about using a fluid velocity of 80% of the gas sonic velocity. This will be the only design criterion for sizing of vacuum heater outlet tubes. Film temperature is not important because of high fluid velocity, which means these tubes won't be limited by coking problems.

2.8. Turndown flow rate

Turndown flow rate can be much lower than the design flow rate (50% of the design rate has been used), because the heat flux also decreases when the VDU unit works below capacity, therefore limiting film temperatures. Turndown flow rate will be limited by the burners' minimum firing rate. Other problems that may limit turndown flow rate are control valve sensibility, development of slug flow (because of the vibrations and pressure swings it causes) or stratified flow (due to dry tube walls) in the tubes and high coking rates in the outlet tubes due to low fluid velocity.

There is no recommended minimum turndown flow rate because each heater has different limitations. The minimum feed rate should be analyzed on a case by case basis. Nevertheless, some engineers say that turndown flow rate should allow at least 70% of the design mass velocity.

2.9. Convection section

A convection section is used to increase the thermal efficiency of a fired heater by recovering as much heat as possible from the flue gas before sending it to the stack or the air pre-heater.
A convection section allows the design engineer to lower the flue gas temperature to around 300-350 ºC, depending on the size of the convection section, the oil inlet temperature and on use of a steam coil. Every new heater is equipped with a convection section in order to recover as much heat as possible. When used, between 30% and 35% of the heat duty is transferred in the convection section and the remaining 65% to 70% in the radiant section of the heater. This will reduce the size of the radiant section as well as the fuel consumption. In existing heaters it can be used to increase capacity or to lower radiant section heat flux.

The convection section has bare tubes and finned or studded tubes. Bare tubes are located in the first 2 or 3 rows and they receive direct radiation from the firebox, for this reason they are called **shield** or **shock tubes**. They are frequently the tubes with the highest heat flux in the heater, which may lead to rapid coking. Although the fluid bulk temperature in these tubes is still low, if care is not taken in their design a very high heat flux can nevertheless increase film temperatures above 460 ºC.

Finned or studded tubes are used above shock tubes where no direct radiation from the firebox reaches them. Their extended surface and therefore higher heat transfer area is used to better heat transfer from the hot flue gases. If the fuel is gas only, 7 or 8 rows of finned tubes are usually used. If the fuel is oil or oil and gas, due to the soot produced, studded tubes are used as well as soot blowers. Since studded tubes have less extended surface per unit area of bare tube, usually 9 or 10 rows are used.

An approach to sizing the convection section is to restrict the maximum convection heat flux from 63 kW/m^2 to 79 kW/m^2 based on bare tube area. [39]

The flue gas mass velocity through the convection section should be 1.0 to 1.5 kg/m^2.s in natural draft heaters and 1.5 to 2.0 kg/ m^2.s in forced draft heaters [30]. Flue gas convection heat transfer coefficients don’t change much, but they depend on flue gas velocity, which may change due to increases or decreases in firing rate or excess air.

2.10. Combustion air

2.10.1. Draft

Draft is the slight negative pressure difference between the inside and the outside of the heater. It is critical to avoid pressurizing the heater’s casing and insure sufficient air flow and adequate air/fuel mixing in the burners.

Natural draft is created by the lower density of the flue gas inside the heater compared to the air in the exterior. Flue gas flows out the stack and combustion air is admitted to the burners in natural draft heaters due to this pressure differential. Since there is some pressure drop through the burners, convection section and stack, sufficient stack height and diameter must be provided to overcome these losses or additional draft must be created using fans to ensure that pressure remains negative throughout the heater. The draft profile of a typical natural draft heater is shown in figure 2.12.

The classification of heaters according to type of draft is the following:

- **Natural draft** – a stack effect induces the combustion air and removes flue gas
- **Forced draft** – uses a forced draft fan or other mechanical means to supply combustion air
- **Induced draft** – uses a fan to remove flue gas and maintain a negative pressure in the heater to induce combustion air
- **Balanced draft** – uses an induced draft fan to remove the flue gas and a forced draft fan to supply combustion air

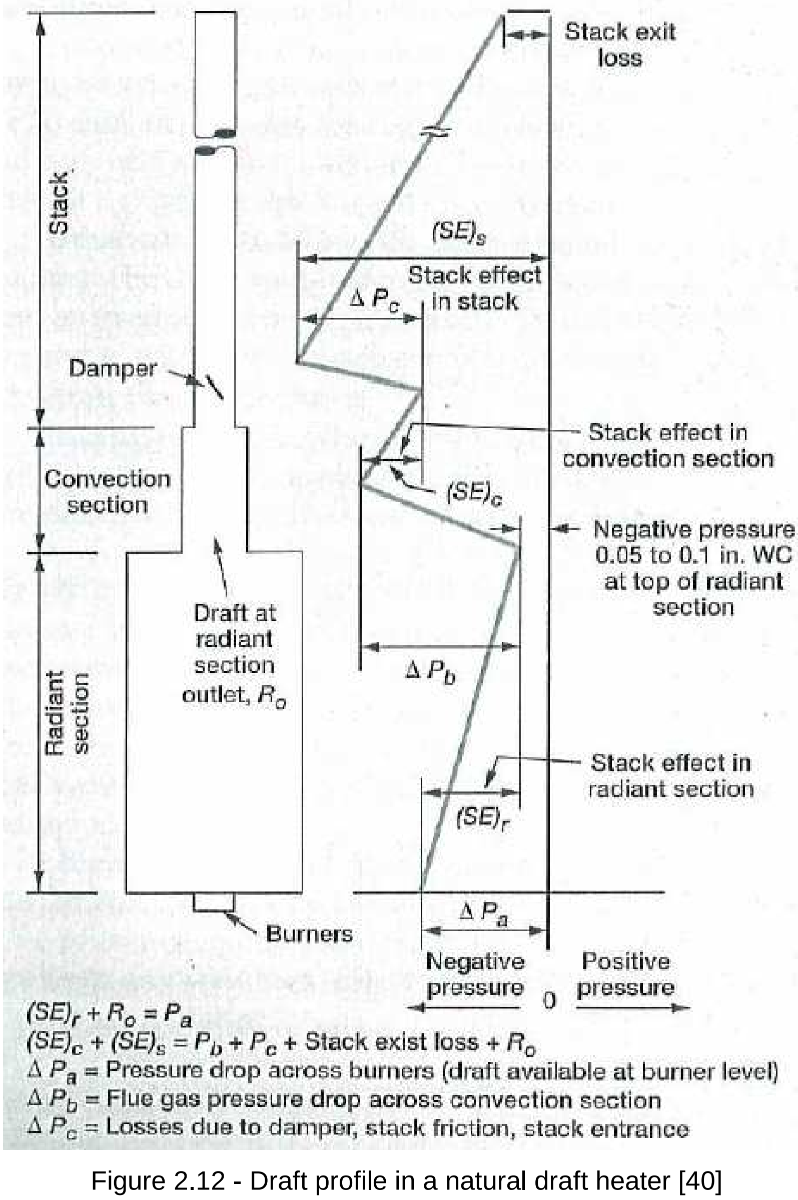


Figure 2.12 - Draft profile in a natural draft heater [40]

The use of an induced draft fan is usually necessary when re-rating or revamping a heater.

In a forced draft system the combustion air is supplied by a centrifugal fan to provide high air velocity for better air/fuel mixing. The stack is still needed to create a negative pressure inside the furnace. A forced draft system is more expensive than a natural draft one but the better mixing of air and fuel allow for the use of less excess air. This means using less fuel which will increase the efficiency of the heater.

Forced draft and induced draft can be used in the same heater and are common with most air pre-heating systems. This is known as balanced draft. It has one great advantage, it enables a very good control of draft and therefore of excess air inside the heater.

The usual measurement point for draft is the heater's arch and the target value is between 1.3 mmH_2O (0.05 inH_2O) and 2.5 mmH_2O (0.1 inH_2O). The first value is recommended because the bigger the draft, the more heat losses we'll have due to air leaks into the heater. This is known as "tramp air".

The stack should be sized for at least 2.5 mmH_2O draft in the heater's arch, at 130% the normal firing rate with maximum ambient temperature and the chosen burners' design excess air. The most economical flue gas velocity is 10 m/s according to UOP [30] and between 8 m/s and 12 m/s according to Garg and Ghosh [39].

One stack should be provided for every 12 meters of convection section tube, or alternatively one flue gas duct leading to the main stack. The temperature at which the flue gas exits the stack is called stack temperature and can be approached to 50 ℃ of the process inlet temperature, although this is not always desirable due to acid mist condensation. For more information on this subject see the next section.

2.10.2. Air pre-heater

As we saw, using a convection section is useful to lower the flue gas down to 300 -350 ºC but this flue gas temperature still represents an important heat loss. In heaters bigger than 10 MW it is attractive to use an air pre-heater to increase the thermal efficiency by recovering as much heat as possible from the flue gas before sending it out the stack. It is not installed in all new heaters because it generally requires a forced draft system and the investment cost of both the pre-heater and the forced draft air system is not always justifiable.

The air pre-heater can lower the flue gas temperature down to 50 ºC but due to the problem of SO_3 condensation when fuel with sulphur is used, flue gas can only be cooled down to 170 – 200 ºC depending on the fuel's sulphur content and the stack wall temperature. The recommended minimum metal temperature for convection coils, fans and duct steel exposed to flue gas is 150 ºC [33]. Nevertheless, an air pre-heater can boost the heater's thermal efficiency by 8% to 18%. Each 20 ºC drop in flue gas temperature roughly represents an additional 1% rise in thermal efficiency.

The heat recovered from the flue gas is used to heat the combustion gas before it is fed to the burners. This increases the firing rate and therefore the heat duty in the firebox. This must be taken into account when sizing the burners and the firebox.

Three types of direct air pre-heater system are used with fired heaters, **regenerative**, **recuperative** and **heat pipe exchanger**. These exchange heat directly with the combustion air. The **regenerative** type uses heated rotating elements and a **recuperative** design uses stationary tubes, plates, or cast iron elements to separate the two heating media. **Although the recuperative type is larger and costlier than the others, it is simpler, requires less maintenance, resists corrosion and needs no power**. [41]

Indirect pre-heaters are also available and they use two exchangers and an intermediate fluid to draw heat from the flue gas and heat the combustion air. This is not very commonly used. The last type of air pre-heater is the external heat source type in which an external heat source is used to heat the combustion air without cooling the fuel gas. This last type of system is usually used to temper very cold air, thus minimizing snow buildup in air ducting and “cold-end corrosion” in downstream air pre-heating systems.

2.10.3. Burners

The function of the burner is to provide adequate mixing of fuel and air. Fuel gas pressure and air draft provide that energy. **Forced draft burners provide better mixing and so have smaller more stable flames than natural draft burners**. For this reason their firing rates – 8 MW to 9.5 MW – are double the ones of natural draft burners – 3.5 MW to 5 MW.

Flames should not be higher than 2/3 of the firebox height according to Shell [9] and ½ the firebox height according to UOP [42], although UOP also says it can be between 1/3 and 2/3 of the firebox length [30]. Fuel gas flame has a more uniform heat distribution than a fuel oil flame. See figure 2.9 above in section 2.4.

Special care should be taken with burners. Flame impingement can result in tube failure and high peak heat flux which are main causes of coking problems. Burners should be in sufficient number and distributed evenly inside the heater in order to avoid high heat flux imbalances between longitudinal sections of the heater’s walls. To this end minimum burner to tube clearances are discussed in section 2.6.3.

It is important to check the burners one by one regularly. Several problems can affect the burners like poor mixing, low fuel pressure, miss aligned, damaged or plugged burner tips, etc (for a more complete list see API Recommended Practice 535). These problems result in fuel waste, heat flux imbalance and even flame impingement. See article [43] on burners.

2.11. Boiling regime

Depending on saturation and bulk and wall temperatures, the boiling regime will change with marked consequences in heat transfer. A fluid with bulk temperature below the saturation temperature can develop subcooled boiling if the film temperature at the wall is above the fluid's saturation temperature. If the bulk temperature is above saturation, temperature the boiling regime will depend on the temperature difference between the fluid bulk and the film temperature at the tube wall. This will determine the amount of vaporization occurring at the tube wall.

When only liquid is present the heat transfer mechanism is forced convection. When the first bubbles appear at the tube wall, the fluid bulk temperature is lower than the liquid saturation temperature. The bubbles will flow towards the center of the tube and condense in the colder liquid, this is called **subcooled** boiling. When the bulk temperature reaches the saturation temperature, **nucleate** boiling develops. In both boiling regimes, heat transfer takes place not only at the tube wall but also in the gas/liquid interface. This increases not only convection in the liquid phase due to the movement of bubbles but also the heat transfer interface area leading to high film heat transfer coefficients. **Nucleate** boiling is characterized by all bubbles having approximately the same size.

If the film temperature at the tube wall increases significantly, either because of a big increased in heat flux or drying of the tube wall due to high vapor fraction (**mist** flow), the boiling regime will develop into **film** boiling. In this boiling regime the tube walls are covered by a film of gas and only the gas phase contacts the tube wall. Heat is transferred by convection through the gas film with very low heat transfer coefficients compared to the other boiling regimes.

Between **nucleate** and **film** boiling regimes a **transition** regime exists, with sections of the tube covered by liquid and others covered by vapour. This regime is unstable and the dry sections are intermittently covered by liquid. The film heat transfer in this regime is lower than in **nucleate** boiling but better than in **film** boiling.

Film boiling develops above a Critical Heat Flux, which changes with heat flux and vaporization fraction. Figure 2.13 shows the critical heat flux variation.

Nucleate boiling is the most effective boiling regime in terms of heat transfer, therefore it is the preferred boiling regime in fired heater radiant section tubes. **Film boiling should be avoided** due to its very low heat transfer coefficient and hence, high coking rate. In the heater's outlet tube **film** boiling may be unavoidable but it won't be a problem if the fluid velocity is kept high, as advised in section 2.7.

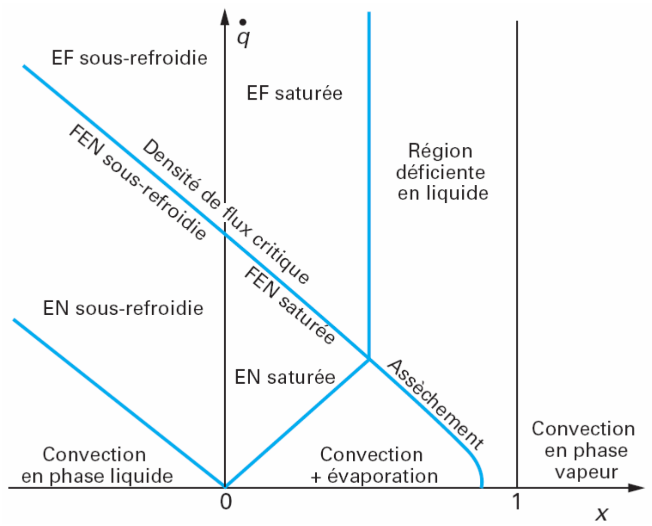


Figure 2.13 - Critical heat flux density map [44]

Key to figure 2.13:

Assèchement: Drying	EN: Nucleate boiling
En phase liquide: In liquid phase	FEN: End of Nucleate boiling
En phase vapeur: In vapor phase	Région déficiente en liquide: Liquid deficient region
Densité de flux critique: Critical heat flux	Saturée: Saturated
EF: Film boiling	Sous-refroidie: Subcooled

2.12. Two-phase flow regime

When only liquid is present the heat transferred to it is received as sensible heat. As the liquid starts to vaporize we enter two-phase flow and most of the heat transferred is received as latent heat. Both heat transfer and pressure drop are affected by the pattern of two-phase flow, which, because of the evaporation of liquid changes along the tubes. Depending upon the orientation of the tubes and the fraction vaporized, several different flow regimes are possible. In horizontal tubes there are additional two-phase flow regimes different from the ones possible in vertical tubes. Some articles describe two-phase flow [45] [46] [47].

2.12.1. Vertical flow

Two-phase flow patterns start with a small fraction of vapor in the form of bubbles that form at the tube surface, which is called **bubble** or **bubbly** flow regime. As more liquid vaporizes the vapor fraction increases and the bubbles coalesce into larger pockets of vapor at the center of the tube separated by plugs of liquid. This creates an intermittent flow pattern, called **plug** or **slug** flow regime, because some slip between the liquid and vapor phases occur, especially at low fluid velocity. This flow regime is unstable and may produce vibrations and pressure swings.

Between **slug** and **annular** flow an intermediary regime may develop called **froth** flow, which corresponds to an intimate mix of vapor and liquid forming a turbulent emulsion.

With increasing vapor fraction the pockets of vapor grow and the gas phase separates from the liquid phase and flows through the center of the tube at high velocity, leaving the liquid flowing at lower velocity near the wall. The film between the wall and the interface narrows as more liquid vaporizes, which reduces the thermal resistance of the film and the wall temperature. The temperature difference between the wall and the interface can be reduced to the point where nucleation at the wall is suppressed, most or all of the vaporization takes place at the interface, this regime is called **annular**.

As the liquid film thins, the interface becomes wavy due to a higher gas phase velocity than the liquid film velocity and small droplets of liquid entrain the gas. As this happens the film may become fractured in some points due to vaporization and the formation of waves. This regime is called **annular flow with entrainment**, **annular mist transition** or just **transition**.

At a certain point the wall becomes dry and all the liquid is in the form of entrained droplets which will continue to vaporize if heat continues to be provided, this is called **mist**, **drop** or **dispersed** flow. In this liquid deficient zone the wall temperature will increase significantly because the wall/vapor thermal resistance is high in these conditions due to low convection heat transfer. Another problem is the contact of liquid droplets with the tube walls, as the walls are very hot the droplets will form coke at very high rates when they hit them.

Figure 2.14 illustrates the different two-phase vertical flow regimes and their effect on heat transfer coefficients.

Annular is the preferred two-phase flow regime but bubbly and even transition can be acceptable. When in **transition** flow, as the regime approaches mist flow the liquid film breaks up and thermal resistance increases, this should also be avoided. **Slug** and **plug** flows must be avoided because they cause vibration problems and pressure swings. **Mist** flow is normal in the heater’s outlet tube due to the high vapor fraction. The high fluid velocity will prevent the deposit of too much coke in the heater’s outlet tube but **mist** flow should be avoided in the rest of the tubes because fluid velocity is too low.

The difference between up-flow and down-flow is the hydrostatic pressure profile that can promote or inhibit vaporization and therefore lead to two-phase flow regime change. This is especially problematic in the top part of vertical tubes with down-flow, where the flow regime becomes unstable after the tube bend. This can cause coking problems in these sections of tube.

Another problem may develop in the tube bends themselves when a significant vapor fraction is present. Due to the centrifugal force liquid flows along the outward curve of the bend leaving the inward curve dry, this can be a cause for high tube wall temperatures on the inside of tube bends. In **mist** flow, erosion can occur in tube bends because of liquid droplets hitting the walls at very high velocity.

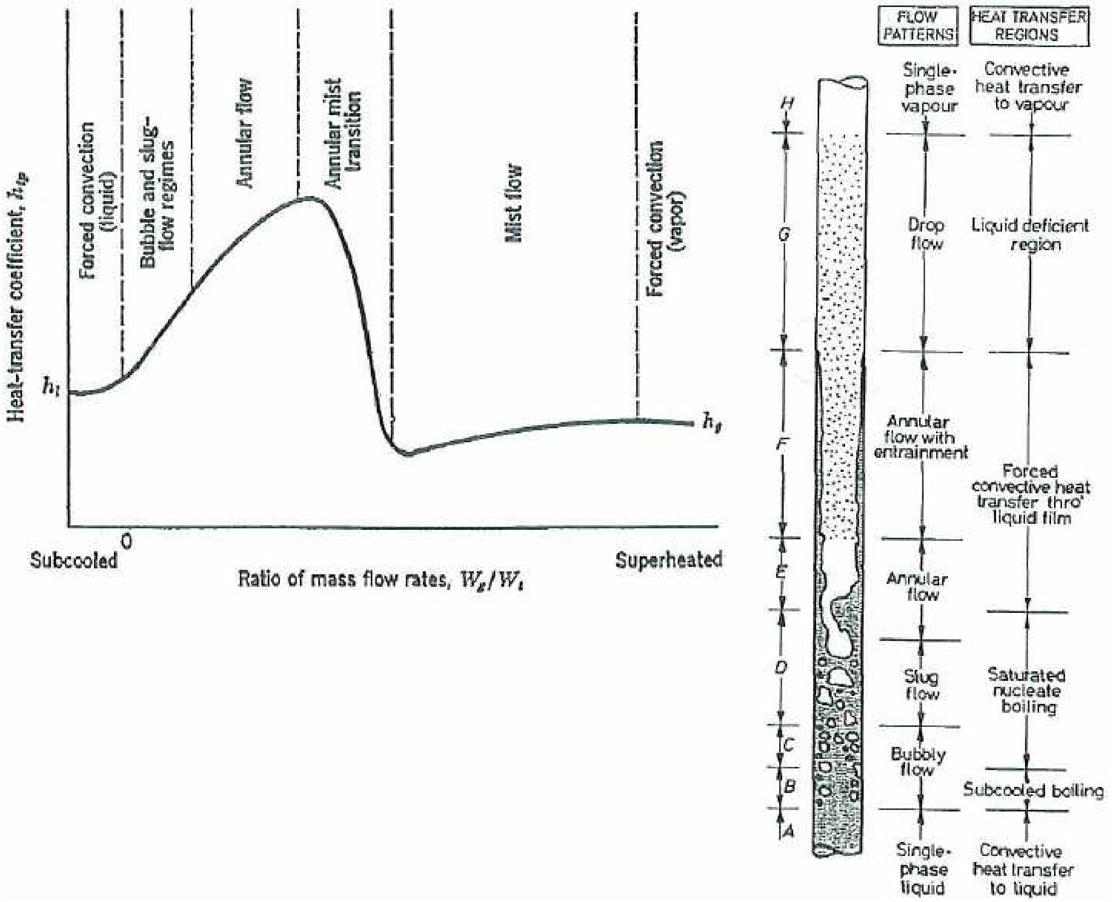


Figure 2.14 - Two-phase vertical flow regimes and their effect on heat transfer coefficients [12]

2.12.2. Horizontal flow

As stated before, there are additional two-phase flow regimes in horizontal tubes due to gravity and the significant difference in density between the liquid and the gas phases. **Bubble**, **plug** and **annular** flow regimes are similar to the ones in vertical tubes with the difference that in horizontal tubes the bubbles or gas phase travel near the top wall of the tube, but tube walls remain completely wet. Also, **mist** flow is similar in both tube orientations, with droplets of liquid dispersed in the gas phase.

But now, at low liquid flows we have additional regimes where a vertical separation of phases occurs drying the top of the tube. **Stratified** flow is characterized by a separate flow with liquid on the bottom and gas flowing on top with a smooth interface. In **wavy**, also called **stratified wavy** flow, waves form in the interface due to increased gas phase velocity. Finally in **slug** flow, large waves, or slugs of liquid extend the whole tube diameter, intermittently wetting the top portion of the tube. As most of the top of the tubes remains dry most of the time it is very hot and this situation will easily lead to coking problems. Figure 2.15 illustrates the two-phase flow regimes in horizontal tubes.

Plug and **slug** flow regimes are unstable and should be avoided because of pressure swings that cause vibrations. We must also avoid **stratified** and **wavy** flow because the tube wall dries, leading to high skin temperatures which will cause high coking rates. As in vertical tubes, **mist** flow should be avoided except in the outlet tubes and for the same reasons. **The recommended two-phase flow regimes for horizontal tubes are the same ones recommended for vertical tubes, annular, bubbly and transition**.

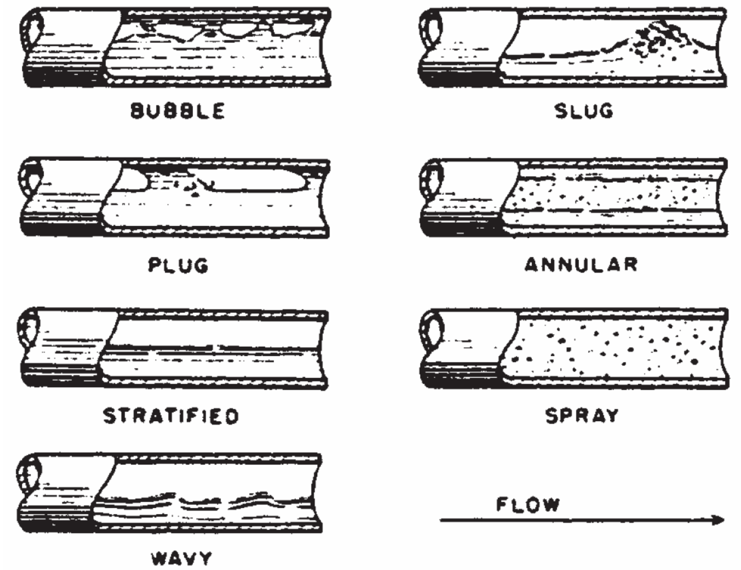


Figure 2.15 - Two-phase horizontal flow regimes [48]

2.13. Conclusions

The conclusions taken from the work done on this chapter are now presented; they were also presented in the text above but their compilation allows for an easier read.

The heater should be sized for the worst operation scenario, predicting probable feed composition and rate changes and also changes in vacuum column cut-point targets. Latter revamps to correct for changed heater operation or increased capacity will cost more than if the heater is built with these changes in mind.

Mass velocity and film temperature are the most important fired heater design parameters because the coking rate is linearly dependent on residence time and exponentially dependent on temperature. The recommended design mass velocity for new vacuum distillation unit fired heaters is 1500 $kg/m^2.s$ and the recommend maximum film temperature allowable in vacuum heaters is 460 ºC.

Balancing the heat flux is the best way to lower oil film temperature. High localized heat flux causes elevated tube metal temperatures even when no coke is present. The current TOTAL average heat flux specification of 26 kW/m^2 for single-fired tubes and 38 kW/m^2 for double-fired tubes should continue to be used as a starting point for heater design, but film temperatures should be checked. Heat flux to all passes must be balanced in order for them to receive the same amount of heat.

Coil steam should be injected before the first tube where fluid bulk temperature reaches the heater's outlet temperature. Adequate coil steam rate has to be evaluated on a case by case basis. When injecting more steam does not promote additional vaporization, more steam should not be used.

Tube equivalent lengths with approximately the same pressure drop should be arranged for all tube passes, so that it does not result in a difference of flow between passes. The pressure drop through each pass should be around 6 bar.

In order to avoid impingement of the flame on heater tubes or refractory the industry has developed minimum standards for burner center line to tube center line clearance, which should be respected.

In the outlet tube the fluid velocity should be 80% of the sonic velocity of the gas phase, which is sufficient to limit coking. In transfer line design, fluid velocity should be limited to around 50% of the gas sonic velocity.

Nucleate boiling is the most effective boiling regime in terms of heat transfer, film boiling should be avoided. Annular is the preferred two-phase flow regime but bubbly and even transition can be acceptable.

There is no recommended minimum turndown flow rate because each heater has different limitations. The minimum feed rate should be analyzed on a case by case basis.

3. Heater operation

Heater operation can have a significant impact on the coking rate because a heater doesn't always operate under the design conditions. In fact a lot of heaters suffer important variations in charge and duty during each cycle [49]. There are also other operating factors that can affect coking, such as flame impingement or bad decoking of the tubes. Additionally, several operational variables have an effect in heater efficiency and we'll discuss them here.

3.1. Charge quality

The composition of the feed charge is an important concern to avoid coking. The asphaltene content (which is a coke precursor) of the charge as well as its tendency to coke can be measured through the peptization state, oxidation level, acidity and asphaltene solubility or flocculation tendency. There are several articles on this subject [8] [14] [15] [20] [22] [23]. Although this may be useful these are not guaranteed methods, we cannot say with certainty if a charge will coke or not. To achieve a good degree of confidence with these methods of prediction, some historic data has to be compiled to be able to evaluate if and how much will the heater coke when processing a given charge.

Oxidized charges usually increase the rate of coke formation, probably due to polymerization of the oxidized compounds [19]. It may be worth considering blanketing residue tanks with an inert atmosphere if it is not common practice.

Another concern is impurities present in the charge. Salts, metals and sulphur all help increasing the coking rate. The influence of heavy metals (including the ones present in the tube alloy) is not very well known but they seem to catalyze the thermal cracking reactions. **Sodium is a well known thermal cracking catalyst**. When in high concentrations in the charge it will increase the coking rate. The increase in coking is squared the increase in sodium content [19]. It is desirable to keep the sodium concentration in the charge below 50 ppm. Above 100 ppm coking will be excessive and therefore unacceptable. Proper desalting of the processed crude is important to limit coking.

Molecules with sulphur atoms seem to be easier to crack, so charges with high sulphur content may suffer more coking. We cannot reduce the sulphur content of the residue but we can blend it with residues with lower sulphur content or we can reduce the heater's outlet temperature when processing high sulphur charges.

3.2. Coil steam

As was referred in section 2.3.2, very high steam feed rates cause elevated pressure drop and may also cause annular flow, with very fast vapor flow in the center of the tube. This leaves the oil creeping along the walls with high residence time aggravating the coking problem. The added pressure drop created by the steam injection increases the boiling temperature of the fluid, in effect increasing film temperatures, as shown by figure 2.6 of section 2.3.2, injecting more than a certain amount of steam has no further effect on the coking rate.

In two-phase flow the main film temperature lowering effect is produced essentially by vaporization, not fluid velocity created by the steam. Fluid velocity has some effect in ameliorating heat transfer coefficients but the increased gas volume due to steam can result in two-phase regime change. Steam has a very low molar mass, so a small quantity will produce a large volume of vapor and may develop annular flow. **If more vaporization does not improve nucleate boiling in the liquid film, the additional liquid residence time may worsen the coking problem instead of helping**.

As a rule, if injecting more steam does not promote additional vaporization it should not be used. There is another downside to more coil steam, it has to be condensed in the vacuum column head, so more motive steam has to be used in the vacuum column ejectors limiting vacuum in the column, which in turn increases pressure in the heater. For these reasons coil steam is usually limited to 1 %wt of the charge rate. In some special situations more steam may help, for example if the charge has a high temperature vaporization curve, but the effect of coil steam injection at high rates should be studied before putting into practice.

3.3. Feed rate

In section 2.1 we saw the effect of mass velocity on the coking rate. We also discussed in section 2.8 that turndown flow rate can be lower than the design flow rate, because heat flux also decreases when the VDU unit works below capacity limiting film temperatures.

In practice the turndown flow rate has limits, if we lower the charge rate too much there will be two-phase flow regime change and undesired flow regimes will develop, such as slug, plug or stratified flow regimes in horizontal tubes, like stratified or stratified wavy. This can cause pressure swings and vibrations in the heater and lead to liquid slip in vertical tubes. When the burners operate at reduced firing rate the heat flux imbalances in the heater may also increase. Although the actual heat fluxes will be lower, the peak to average heat flux ratio in the heater may increase, which will cause problems due to the reduced mass velocity.

Another caution with feed rate is to avoid sudden decreases that can lead to rapid coke deposition. If less feed passes through the tubes while at the same firing rate, heat flux will be too high and the outlet temperature and vaporization will increase. This means elevated temperatures and vaporization in the heater with the consequent film temperature increase and possible two-phase flow regime change, which may result in very high coking rates. Since refractory walls store an important amount of heat, it is not enough to reduce fuel to the burners. We must wait for the walls to cool to the temperature corresponding to the new firing rate. **When reducing unit throughput the firing rate should always be reduced before the feed rate.**

Sometimes this cannot be avoided as when there is a problem with the feed pumps, a valve, a flow-meter or even a burst pipe. For this reason, **when the feed is cut or significantly decreased, high pressure purge steam must be automatically opened into the heater coil inlets to clear the oil from the tubes and keep them cool until the feed can be restored or the heater is shutdown.**

If flow is restored too soon while the tubes are still very hot, the first liquid to run through the tubes will almost instantly turn into coke. Time must be given for the tubes to cool to an adequate temperature, that is, down to normal film temperatures.

In some start-up procedures low feed flow may be inescapable, in such periods steam injection, but preferably recycle or adding another hydrocarbon current should be considered to maintain minimum flow in every pass. **Stable and equal pass flows are important at start-up as well as avoiding long start-up periods to prevent initial coke lay-down**.

3.4. Cracked gas and coking

Cracked gas (vacuum ejector incondensables) can give a measure of the coking rate. If we can measure the vacuum column off gas (which is essentially cracked gas), we can infer the coking rate. Since coke and cracked gas are both formed when thermal cracking occurs, the more coke, the more cracked gas.

When using this other reasons for high vacuum ejector incondensable rates should be checked, such as high vacuum distillation column bottoms level (high residence time will lead to high thermal cracking) and air leaks into the vacuum tower.

However, even with reliable and accurate cracked gas measurements, we can neither calculate the exact coking rate nor where coke is forming inside the heater. What we can learn from cracked gas measures is the severity of coking. If the cracked gas produced is less than 0.05% wt of the feed, the coking rate is slow. If it is higher than 0.2% - 0.3% the coking rate is probably too high and we'll have a short run length.

3.5. Decoking

If a heater is not properly decoked deposits will remain in the tubes. These deposits will serve as coke initiators either by catalyzing thermal cracking or just by increasing wall temperatures. A good method to evaluate the decoking is to compare start of run skin temperatures with the first run of the heater when tubes where new.

Several decoking methods are used: **spalling**, **steam/air decoking**, **pigging**, **steel shot**, **sandjet**, **chemical cleaning** and even **tube replacement**. **Steam/air decoking** and **pigging**, are the most widely used decoking methods in vacuum and visbreaker unit heaters.

Steel shot and **sandjet** use, respectively small steel balls or sand, to erode coke from the tube walls. This has the disadvantage of also eroding the tubes.

Tube replacement is obviously an expensive alternative and is reserved for the replacement of damaged tubes.

Spalling consists of passing steam through the tubes while manually firing the furnace to raise and drop coil outlet temperature in a programmed sequence. The temperature differences cause the coke to crack due to thermal stress (expansion/contraction cycles) so that it loosens from the tube walls and the steam can tear it off. This procedure does not completely remove coke, it only removes large pieces of coke not the whole layer and it works only with friable coke, so it is mainly used in coker units to extend run length. In vacuum or visbreaking units a filter is generally used at the end of each coil to collect the coke.

Steam/air decoking is similar to **spalling** but air is injected to burn the remaining coke. Steam is continuously passed through the tubes to protect them and control tube temperatures. Coke removal is usually done by a combination of **spalling** and burning to decrease the quantity of coke that must be removed by burning. This process has the risk of damaging the tubes while burning coke if caution is not taken to ensure tube metal temperatures do not pass their limit metal temperature. This method has an advantage, it can be performed on-line, one coil at a time, without shutting down the unit.

Chemical cleaning is used to remove salt deposits and consists of circulating an inhibited acid through the coil until all inorganic deposits have been softened and removed. This is usually followed by water washing to flush all deposits from the tubes.

Pigging is a more recent method of decoking, it is a mechanical cleaning method and is widely employed. It uses a plastic foam pig with metal studs attached to the body driven by water pressure. Pigs are used in incremental sizes to avoid getting stuck. They can travel both ways in a tube allowing for its release if the pig gets stuck and also to clean or polish only a specific section of tube. Advantages include no use of heat and erosion can be controlled by monitoring the water, so there are no risks to the tubes and it can clean not only coke but also inorganic deposits. The tube is polished at the end of the pigging process resulting in a smooth tube surface as in new tubes. It can also provide the operators with a report on coke location and approximate thickness, obtained from the pig sizes used and pressure drop monitoring during cleaning. It can also provide coke samples for laboratory analysis. Heater connections (thermocouples and steam injection points) must be prepared for pigging or removed before the procedure. Return bends should be used throughout the coil instead of square headers and places for pig launch and recovery must be installed.

Pigging is reported to be faster and more effective than other decoking methods, as well as a cost effective, safe and more environmentally friendly alternative to steam and air decoking [50]. It doesn't expose the tubes to high temperatures and removes inorganic deposits that cannot be removed by **steam/air decoking** or **spalling**. Inspection of the tubes after pigging in several heaters shows no erosion problems. Many heaters show lower skin temperatures at start of run after pigging than after other decoking methods.

3.6. External tube cleaning

Not only internal fouling, but also external fouling affects heater tubes. As with coke this external fouling layer poses an additional resistance to heat transfer. In this case tube metal temperatures are not increased but increased firing is necessary to maintain heater outlet temperature. This results in wasted fuel, higher bridgewall temperature and flame impingement in some situations. **An externally fouled firebox can waste up to 10% of the fuel** because of the higher flue gas temperature necessary to compensate for additional heat transfer resistance and maintain the same heat flux to the tubes. If clean areas exist amid generally fouled tubes, these clean surfaces will experience high heat flux that may translate into high tube wall temperatures.

This external fouling layer is composed mostly of ash from fuel burning but can also have corrosion products due to flame impingement on the tubes or sodium/vanadium corrosion when heavy oil is fired. External fouling is worst when fuel oil is fired but even fuel gas can produce some ash. External deposits come from FCC catalyst fines, carbon particles due to incomplete combustion and in the case of fuel oil, salts and heavy metals.

When external fouling is considered excessive there are methods for on-line cleaning. Tubes can be blasted with sand or crushed walnut shells or chemically cleaned. These methods have very quick payback and cause minimal disruption to heater operation. Access to tubes via portholes and sight glasses typically limits the ability to clean 100% of tube surface.

Chemical cleaning uses ammonium, potassium or magnesium nitrates together with inhibitors, catalysts and a pH-control colorant in an aqueous solution carried by an air spray. The chemical reacts with fouling material causing, oxidation and subsequent burning of the partially disintegrated deposits and protects surfaces from sulphuric acid and vanadium corrosion. Additional mechanical action by air jet removes the entrained deposits. Heater refractory fibbers can be damaged by the high velocity air jet but the risk can be minimized. This method is used on oil and gas fired heaters.

Sand blasting is used on gas fired heaters, which tend to foul from the formation of metal oxide scale on the exposed surface of the heater tubes due to high temperatures. If sand is used erosion of the tubes is a concern and sand will remain inside the heater. Using crushed walnut shells we avoid these problems with no increase in emissions. The danger of this method is the same as chemical cleaning, heater refractory fibbers can be damaged by the high velocity blasting media.

The effect on the bridgewall temperature can go from a 50 ºC to a 120 ºC reduction. Sometimes better draft is also noticed after cleaning. **Two radiant section cleaning operations per year are normally suggested in oil fired heaters**.

Convection section tubes, especially those with extended surface in oil fired heaters, are prone to external fouling by soot which reduces draft and heat transfer due to a reduction in effective surface. The same methods described above for cleaning external fouling in radiant section tubes are used for convection section cleaning but they are usually used employed as off-stream methods, as is water jet blasting.

On-stream mechanical cleaning devices are used regularly to blow away the soot, sometimes additives are also used in conjunction to soften the soot deposits. These mechanical devices are called soot blowers and use steam or compressed air to blow soot from the tubes' extended surfaces but they are ineffective on fused deposits. That is why chemical addictives and off-stream cleaning methods are necessary.

There are two types of soot blowers. Rotary soot blowers use a multi-holed lance that remains in position in the flue gas passage. When activated the lance rotates with steam or air spraying from the holes to dislodge deposits. This type of lance should not be used in high temperature and corrosive areas. It can clean up to three rows from the lance plane in a maximum radius of 1 m.
Retractable soot blowers have two holes at the end of the lance and are normally retracted from the flue gas passage. When in service the lance rotates and travels through the flue gas passage and returns. This type of lance is used in high temperature and corrosive areas. It can clean up to four rows from the lance plane in a maximum radius of 1.2 m.

Steam pressure should be at least 10 barg but it is limited to 20 barg, higher pressure may cause tube erosion. A minimum clear depth of 0.5 m should be provided to allow the blowers to be effective. Also the effective radius of cleaning should be known when determining the number of soot blowers necessary. Soot blowers are normally used in 8 hour cycles.

3.7. Excess Air

Since perfect mixing of fuel and air is not possible, **excess air is needed to ensure complete fuel combustion**. It is expressed as a percentage of the theoretical quantity of air required for perfect combustion, known as stoichiometric quantity.

Forced draft burners have a higher pressure drop which promotes better air/fuel mixing, therefore requiring less excess air. The type of fuel is also important, gas burners require less excess air than oil or oil/gas burners to achieve complete combustion.

Burners are designed to operate at a certain excess air percentage to provide sufficient air/fuel mixing. This excess air allows them to operate at their maximum efficiency. This specification can be found on the heater's or burner's documentation. Air registers should be used to control the excess air. Table 3.1 shows the intervals of excess air values used in fired heaters.

Table 3.1 - Excess air by draft and fuel type[2]

	Fuel Gas		**Fuel Oil**	
	All values	Most common values	All values	Most common values
Natural Draft	15% - 20%	15% - 20%	25% - 40%	25% - 30%
Forced Draft	5% - 15%	10%	10% - 25%	15% - 20%

[2] Bibliographique sources: [7] [30] [36] [39] [41] [63]

To maximize heat losses, excess air can be kept below 5% by using high intensity burners or forced draft burners with air preheating, [43] which promote good air/fuel mix.

For maximum combustion efficiency, it is important to establish equal air distribution to all the burners, especially when the excess air is limited to less than 5%. One way of accomplishing this is via directional splitters. Where to locate the splitters can be determined by means of flow modeling in an accurately scaled transparent model of the ducts and windbox. [51]

In some situations it may be desirable to increase excess air even at the expense of burning more fuel. Higher excess air transfers duty from the radiant section to the convection section because the higher flue gas flow, lower flame and bridgewall temperatures reduce radiant heat flux. This also reduces flame size and therefore flame impingement.

3.8. Excess O_2

It isn't easy to measure excess air so O_2 is usually measured at the firebox exit to evaluate the efficiency of combustion. Excess O_2 is a function not only of excess air but also of the fuel carbon content, so if the fuel composition varies excess air must be calculated from the excess O_2 and used as the control variable.

Operating a process heater simply to achieve a minimum excess oxygen target in the flue gas can waste a great deal of energy. The proper way to adjust O_2 in a heater is to target for the point of absolute combustion. The point of absolute combustion is defined as the air rate that maximizes heat recovery to the process. That is, either a decrease or an increase in the combustion air supply will reduce heat absorbed in the heater.

Some of the objectives in the operation of a fired heater are to avoid excessive heat flux in the firebox, maintain a small negative pressure below the bottom row of shock tubes and obtain complete combustion of the fuel in the firebox with minimum oxygen in the flue gas. Regardless of the flue-gas oxygen content, however, the point of absolute combustion is the combustion air rate that maximizes process-side heat absorption for a given amount of fuel.

The target for excess O_2 will be the point of absolute combustion, provided that all fuel is combusted at this level of excess air.

The oxygen target required to achieve the point of absolute combustion may be 2%, or it may be 6%, depending on the operating characteristics of the heater. Often, energy is wasted because operators and engineers fail to understand that excess oxygen of 2% to 3% in the flue gas is not a good target for all heaters.

The O_2 content of flue gas is not always a good measure of combustion. If the flue gas analyzer is located in the stack tramp air will void the O_2 measures. For this reason the flue gas analyzer should be located at the heater's arch before flue gas enters the convection section.

The goal is to minimize excess O_2 while not allowing fuel to go unconsumed [52]. CO analyzers are often used to detect incompletely burned fuel and the goal is usually to keep it below 150 ppm or some lower target. Excess O_2 is controlled to stay as low as possible without exceeding the CO limit, which is usually 2% O_2 or less at the arch.

3.9. Flame impingement

Flame impingement causes tube coking and may lead to rapid tube failure. It must therefore be avoided at all costs.

Flame dimensions are important to avoid flame impingement and this is to be taken into account when choosing the burners. Flame pattern and shape can be altered by changing the oil spray angle and tip position. A clear spacing of at least 900-mm must be maintained between the tube face and the outer edge of the flame. Flame height should be restricted to 50% of the firebox height at maximum liberation.

The flame pattern improves considerably in forced draft burners when preheated air is used. Although it is difficult to evolve a formula for flame dimensions, most manufacturers predict a flame length of 1.5 to 2.0 meters/MM kcal for forced draft burners. The flame diameter and to a certain extent flame length, is guided by the shape of the burner quarl (or block). Most burner vendors estimate the width of flame as 2 to 2.5 times the burner quarl diameter. [43]

In order to avoid impingement of the flame on heater tubes or refractory the industry has developed minimum standards for burner center line to tube center line clearance. These clearances are shown in figure A.1 in the annex section. Because of their complexity, process heaters will not be firing at optimum conditions all the time. That is why **companies with extensive heater operating experience recognize that the API minimum clearances offer inadequate protection** against common events such as burner fouling, operating upsets, rapid fuel and process changes, etc. Figures A.2, A.3 and A.4 in the annex section show recommendations by UOP and engineers with extensive experience in heater design.

Special care should be taken with burners to avoid flame impingement. Burners should be in sufficient number and distributed evenly inside the heater in order to avoid high heat flux imbalances between sections of the heater's walls. Also it is important to check the burners one by one regularly. Several problems can affect the burners like poor mixing, low fuel pressure, miss aligned, damaged or plugged burner tips or others (for a more complete list see API Recommended Practice 535). These problems result in fuel waste, heat flux imbalance and even flame impingement.

3.10. Heater control

Control strategies can have an impact on heater efficiency and even in coking [53]. Adequate draft and excess air control can improve efficiency and save fuel, pass flow imbalance is responsible in some heaters for increasing coke [40].

3.10.1. Pass balance control

Some heaters do not have individual pass flow control. If fluid is not controlled to each pass there will be flow rate imbalance between passes, which will lead to high coking rates in the tubes with lower flow rates. Coke deposits will increase the pressure drop in these tubes, decreasing even further the flow rate and depositing more coke on the tube walls. This process will make skin temperatures rise rapidly, once they reach the tubes' metallurgic limit the heater will have to be decoked. Therefore **individual pass flow control must be used in fired heaters in coking service**.

There are two usual ways of controlling individual flow to the passes, one is to control flow to each pass in order to obtain the same outlet temperature and the other is to use the same flow rate in all passes.

The same outlet temperature for each pass can be obtained by increase of flow to a pass with a higher outlet temperature. Because at the same firing duty heat flux in the heater is more or less constant, it doesn't matter much what is the flow rate in a pass, all passes will receive about the same amount of heat independent of flow rate. This means that when operating to equalize outlet temperatures, the controller will increase flow to a pass with a higher outlet temperature.

Outlet temperature balance is fine while coils are clean, once coke starts to form in a pass it acts as insulation. This increases skin temperature and the tube reradiates more of the received heat. While this will not result in a big heat duty difference, it should be enough to lower the pass outlet temperature and lead the controller to lower flow rate to the pass. This will result in reduced mass velocity aggravating the coking problem.

The point is that **operating to equalize pass outlet temperatures with large flow disparities will likely cause accelerated in-tube coking**.

Every Standard, Design Recommendation, Memo and scientific article on this subject, either by TOTAL or a third party, states that **pass flow rates should be kept relatively equal**. Some operators recommend that the difference in flow between passes not exceed 1% of the total charge flow, although others relax that limit to 5% of the lowest pass flow.

At times it may be necessary to have unbalanced pass flows to correct for an unbalanced firing situation, for example due to off-line burners. This should not be allowed to continue for long. Firing imbalance must be corrected as soon as possible. On end of run conditions pass flow may be selectively adjusted to control tube skin temperatures and therefore provide a small run extension.

3.10.2. Draft control

Draft should be controlled to ensure proper excess air and minimize air leakage into the heater. It should also ensure that the whole heater is under negative pressure to avoid deterioration of the heater's structure by flame impingement and prevent the flame or hot gases from exiting the heater from any point other then the stack.

The draft profile inside the heater can be adjusted with both the stack damper and the burner's air registers. Using only one of them will not permit to achieve the best point of operation.

The damper is used to control draft in the heater in order to avoid positive pressure and minimize air leakage into the convection section. Air registers should complement the damper in order to achieve the desired excess air target. Both the damper and the air registers affect draft, which in turn affects air flow into natural draft burners. So damper and air registers are incrementally adjusted by the control system until optimum draft and excess air are achieved.

The controller must use the opening of the damper and air registers to control the draft while maintaining the desired air excess. Failing to keep the correct air excess will result in wasting fuel. Unfortunately air registers are usually only automated in furnaces with forced draft burners.

3.10.3. Firing control

Rapid and large changes in fuel input during start-up or after temporary reductions in feed rate should be avoided to prevent high heat flux/low process flow incidents that are sure to accelerate coking.

Whenever possible reduce fuel input concurrently with feed rate reduction to avoid short term low flow, high heat flux events. Feed-forward firing control based on feed rate can handle such events automatically and should therefore be considered.

Cyclic firing patterns and significant fuel input overshoot will cause high heat flux incidents and may also result in flame impingement. This requires fuel and process flow controls to be kept properly tuned. Feed-forward firing control can also help reduce this type of incidents.

In theory, feedback control fixes everything. In practice, there are problems. Fired heaters have a considerable response lag to changes in either the firing rate or process inlet conditions. There will be transient dips or bumps whenever a significant change occurs on any process, fuel, or air variable. Feed-forward is the technique of preventing the problem before it occurs.

Fuel gas can vary significantly in composition and therefore in heating value, especially if the hydrogen content suffers significant variations. The fuel must be controlled to provide the exact amount of energy that the process requires. This implies that the heat content of the fuel as well as its flow rate must be known. If the fuel pressure, temperature and composition are constant, a simple flow meter is sufficient. If significant variations are expected, more complex feed-forward techniques can be applied. In such cases a fuel specific gravity analyser or a fuel calorimeter can be used to measure the fuel's approximate heating value and use this value in the heaters firing control system.

Calorimeters function by actually burning a small amount of fuel that has been metered through a fixed restriction orifice. In this manner the effect of changing density on the flow meter and the control valve is mimicked. They work very well with mixed hydrocarbon fuel gas. The assumption is that there is a proportional relationship between heat content and oxygen demand. This assumption is reasonably accurate for hydrocarbons but is not reliable if there are large, variable proportions of hydrogen in the fuel. In such cases, more complex analyzers are needed to provide an exact oxygen demand value which can be used to adjust the ratio of the air controller.

When there are large, variable proportions of hydrogen in the fuel a chromatograph can be used to measure the composition. From this density, heating value and oxygen demand can be calculated. Combining these calculations with pressure, temperature and flow readings result in an exact value of heat input per unit of time. The result of this arrangement is that any changes in fuel gas condition will be corrected before any change in the process temperature occurs.

3.11. Heater monitoring

Visual inspection, especially by an experienced operator, can give valuable information about fired heater operation that is not available from instrumentation. It may also give some indication of instrument malfunction that can otherwise be ignored.

Regardless of oxygen measurements, the appearance of the firebox must always be taken into account. A combustion zone that looks bright and clear has excess oxygen. Long, licking, yellow, smoky flames and a hazy firebox indicate low oxygen. Yellow flames in themselves, however, are fine. A gas burner may be perfectly adjusted and still have a short, bright, yellow flame. An optimized firebox should look hazy because the fuel is groping for the last oxygen in the firebox.

Attention must be paid to flame size and stability and flame impingement must be promptly corrected. Flame impingement is often caused by poor burner condition (tip erosion, size, coking, plugging, misalignment, etc.). Operators should regularly observe flame patterns and fuel pressure at the burner to judge when burner cleaning or repair is necessary and which burners are affected. Burners should be thoroughly checked every scheduled shutdown.

Heater tubes and refractory will glow different colours, depending on their exterior temperature, see table 3.2 for a colour table. A freshly cleaned heater fired at a moderate rate will have black-appearing tubes. In contrast, the hangers (i.e., the brackets supporting the tubes) will be glowing red. The process fluid keeps the tubes, but not the hangers, cool. A moderately hot temperature is indicated by a dark red colour. As this colour brightens, the tube is becoming hotter.

A hot spot on a tube is usually caused by a partial loss of flow through one pass of a multi-pass heater. In most refinery services, a sudden reduction in flow will cause tubes to overheat. The low velocity combined with high tube-wall metal temperature results in localized coke lay-down. The layer of coke insulates the tube wall from the cooling effects of the process fluid, causing the insulated portion of the tube to overheat. Typically, **hot spots due to coke look like silver streaks or spots**.

An infrared thermograph survey is one way of obtaining a more accurate measurement of temperatures in the heater while on-stream [54]. It allows the identification of hot-spots in the tubes and refractory, showing flame impingement, high heat flux areas and coked spots inside the tubes. Skin temperatures obtained through this way can be compared to readings by skin thermocouples to evaluate their error.

Table 3.2 - Approximate heater temperature based on colour of the material [55] [3]

Colour	Temperature, ºC
Dark blood red or black red	550
Dark red	600
Dark cherry	650
Bright cherry	750
Light red	850
Orange	900
Light orange	950
Yellow	1000
Light Yellow	1100
White	1200
Dazzling white	1500

In oil-fired heaters, ash accumulates on the radiant tubes. These ash deposits may glow red, while the tube wall metal is relatively cool. Care must be taken to identify ash or corrosion product deposits in the tubes' outside walls, these pose a resistance to heat transfer and will therefore be hotter than the actual tube outside wall. This may make difficult the identification of hot-spots due to coke and high heat flux by thermograph survey.

Skin thermocouples are used to measure the tube metal temperature, which gives the operator a measure of coke accumulation from the rise of skin temperatures and indicates the need for decoking when tube metal temperatures near their limit.

In vacuum heaters, 3 to 12 skin thermocouples are used per pass. In each tube equipped with skin thermocouples, 2 or 3 are placed along the tube to measure longitudinal heat flux imbalance. In each coil, skin thermocouples are always placed on the outlet tube. Other usual places are shock tubes, the first section of radiant tubes, tubes nearer to the flame, the tube before the steam injection point and the tube where the wall is expected to dry, if that is expected. If infrared thermograph scans reveal consistently the same tubes with high skin temperatures, these should also be equipped.

3.12. Conclusions

Heater operation can have a significant impact on the coking rate because a heater doesn't always operate under the design conditions.

[3] Colour evaluation is very subjective; a handheld pyrometer will help measure spot temperatures

Charge quality affects the coking rate. Residues with high asphaltene content will produce more coke as will residues with little asphaltene solubility. Oxidized charges usually increase the rate of coke formation and sodium is a well known thermal cracking catalyst.

When reducing unit throughput the firing rate should always be reduced before the feed rate to avoid high heat flux incidents. When the feed is cut or significantly decreased, high pressure purge steam must be automatically opened into the heater coil inlets to clear the oil from the tubes and keep them cool until the feed can be restored or the heater is shutdown. Stable and equal pass flows are important at start-up as well as avoiding long start-up periods to prevent initial coke lay-down.

Rapid and large changes in fuel input during start-up or after temporary reductions in feed rate should be avoided to prevent high heat flux/low process flow incidents that are sure to accelerate coking.

Individual pass flow control must be used in fired heaters in coking service to ensure the same feed rate to each pass. Operating to equalize pass outlet temperatures with large flow disparities will likely cause accelerated in-tube coking.

Pigging is reported to be faster and more effective than other decoking methods, as well as a cost effective, safe and more environmentally friendly alternative to steam and air decoking.

An externally fouled firebox can waste up to 10% of the fuel, so two radiant section cleaning operations per year are normally suggested in oil fired heaters.

Burners are designed to operate at a certain excess air percentage to provide sufficient air/fuel mixing, therefore excess air is needed to ensure complete fuel combustion. It isn't easy to measure excess air so O_2 is usually measured at the firebox exit to evaluate the efficiency of combustion. The target for excess O_2 will be the point of absolute combustion, provided that all fuel is combusted at this level of excess air. Proper fuel combustion can be evaluated through a CO analyzer.

Draft should be controlled to ensure proper excess air and minimize air leakage into the heater. The draft control point is the arch, below the shock tubes and the target is usually 0.05 inH_2O. The draft profile inside the heater can be adjusted with both the stack damper and the burner's air registers. Using only one of them will not permit to achieve the best point of operation.

4. Vacuum heater operation at TOTAL

TOTAL operates several refineries in Western Europe as well as in the United States and has interests in refineries all over the world. This represents dozens of vacuum distillation units, but information on these units is not widely available inside the group. A request has to me made to each refinery, detailing the desired information and wait for the unit engineers to reply, which depends on their current workload.

The information presented here was compiled from several sources. The main source was a Benchmarking study of TOTAL Vacuum Distillation Units [32] that sought to compare the efficiency of VDU units but also presented some heater data. Another important source was a memo that focused on four VDUs from the Grandpuits, Feyzin, Flanders and Provence refineries [31]. Additional information was collected on the Feyzin and Leuna vacuum heaters, which were simulated with FRNC-5 yielding further data.

4.1. Compiled VDU data

The benchmarking study [32] is very specific in scope since it was mostly intended to compare energy efficiency. The memo [31] however was written as an evaluation of the revamp potential of one of the units. Its focus is the heater and pre-heat train but it was extended to the column, transfer line and vacuum system.

This memo's data allowed for the calculation of mass and fluid velocities in the inlet tubes. Simulations of the Leuna and Feyzin vacuum heaters run with FRNC-5 yielded further results. The data that resulted from calculations and simulation is presented in table 4.1.

Table 4.1 - Calculated vacuum heater data

	Feed rate	**Inlet tube mass velocity**	**Heat flux**	**Inlet Tube Velocity**	**Vaporization**	**Run length**
	t/h	kg/ m^2 s	kW/ m^2 h	m/s	%	months
DGS	520	1700	-	5,1	-	60
FZN	230	2690	37.5	3,4	64 – 72	60
GPS	260	950	31.4	1,2	50 – 70	35
RF	230	1237	34.9	1,55	60 – 75	24
RP	300	1105	32.5	1,4	50 – 75	14
TRM	637	1200	-	4,7	60	24

VDU transfer line data is scarce in the available documents. There is pressure and temperature data from the Milford Haven VDU transfer line presented in table 4.2, this differs from the data presented in table A.1 in the annex section because table 4.2 refers to a particularly difficult run. Transfer line temperature drop data from other VDUs is shown in Table 4.3.

Table 4.2 - Milford Haven transfer line data [32]

Pin	150	mmHg
Tin	413	℃
Pout/flash	35	mmHg
Tout/flash	399	℃

Table 4.3 - Transfer lines data[4]

	Leuna	**Feyzin**	**Grandpuits**	**Provence**
Temperature drop	17 ºC	10 ℃	10 – 15 ℃	15 – 20 ℃
Length	35 m	20 m	-	45 m

Table A.1 in the annex section, sums up the available information considered most relevant to compare VDU unit performance and operation. The values shown correspond to average operation except when more than one value is presented.

Some data show two values, which correspond to two different types of run when a slash separates the numbers (some heaters process alternately, high and low sulfur crude mixtures, in runs using different operating conditions) and two values separated by a hyphen indicate an interval of operating conditions. When three values are shown for the same variable, the values separated by a hyphen are the extremes of the interval of operating conditions and the value separated by a slash is the average value.

Empty cells with grey shading represent data not available. The Milford Haven and Port-Arthur VDUs are dry units; steam is used, neither in the heater nor in the column bottom. The Normandy D12 unit is still in project phase, the values presented are design values and the run length is an estimate. The Flanders VDU heater suffered a revamp with the installation of tube inserts which is expected to extend the run length from one year to two years. The last two rows show average values and the ones considered as the best for improved VDU operation. Two smaller tables under table A.1 present heater conditions of VDUs with more than one heater, LOR – VDU2 and Antwerp.

[4] The higher temperature drop in the Provence VDU transfer line is attributed to critical flow. Values were obtained from FRNC-5 simulations and reference [31]

4.2. VDU data comparison with literature values

Starting with table 4.1 we can see that run length, and therefore the coking rate, have a very good correlation with in-tube mass velocity. This correlation also depends on heat flux. This agrees with what was stated in section 2.1 of this work.

Tables 4.2 and 4.3 refer to VDU transfer lines. We can see that a 10 ℃ to 15 ℃ is the usual temperature drop in a transfer line. As for transfer line pressure drop there is only one value available, which is a low pressure drop according to section 2.7.

As for table A.1 in the annex section, some of the values are given as intervals of operation. These are more difficult to compare since we don't know which point of the interval corresponds to other values given for that VDU.

The heaters' outlet temperatures presented vary substantially. This has to do with the severity of operation. Most heaters operate at high severity, with outlet temperatures above 410 ℃, while a few have outlet temperatures below 400 ℃. The maximum outlet temperature is 430 ℃, which is usually assumed as the maximum allowable outlet temperature in vacuum service. [6]

Vaporization is high, between 50% and 75% in all the VDUs for which we have data except for the Roma VDU. This VDU has a vaporized fraction in the flash zone that is almost half the value for the others. We can also see that this heater has a low outlet temperature, which probably explains the low vaporization, but despite low temperature the heater still has short run lengths due to coke. This suggests that low velocity in the heater's outlet tube leads to coking problems.

Excess O_2 is a function of excess air and this is the variable that is measured. A 3% to 4% excess O_2 corresponds roughly to a 15% to 20% excess air, so the values shown correspond to the normal excess air values presented in table 3.1.

Energy efficiency was the main objective of the benchmarking study [32]. Most of the heaters have a standard efficiency of around 80%, give or take 5%. Since the maximum efficiency attainable is 92%, this probably means not enough heat is being recovered from the flue gas. This can be confirmed by the flue gas temperature, which is indeed higher in the heaters with lower efficiencies.

The VGO cut-point is a measure of vacuum column distillate yield. High VGO cut-points represent high yields. To achieve high VGO cut-points, high heater outlet temperatures and low flash zone pressures are necessary, as we can see by looking at the values in table A.1

in the annex section. Flash zone pressures are around the expected values, between 65 mbar and 135 mbar. As for flash zone temperatures, there are some values higher than the expected 400 ℃ but this only means the refiner is exploring the VDU’s potential for higher cut-points.

There is a linear tendency between VGO cut-point and both flash zone temperature and pressure. These two, the efficiency of the separation in the vacuum column and the composition of the residue being processed define the VGO cut-point.

Feed rate and heat duty give an idea of the size of the VDU but are not relevant to this study.

Feed specific gravity is a measure of the residue’s heavy components content. This can vary a lot even in the same VDU, depending on the crude oil fed to the atmospheric unit and also the atmospheric unit’s separation efficiency. A feed specific gravity near or above one indicates a very heavy residue.

The differences between run lengths are enormous and the reasons are not all evident and can have different sources. All run lengths are defined either by maintenance stops or decoking stops, but all short VDU run lengths are caused by coking. The causes of coking can be different and several at the same time, as discussed in section 1.3.

5. Revamp considerations

Process heaters can be limited by heat release, draft, or heat absorption. The process engineer seeking to expand furnace capacity must first determine which of these problems is limiting. Revamps can be aimed at increasing fired heater capacity, severity, thermal efficiency, run length or more than one of the above [37] [41] [56] [57] [58]. Some of the possible modifications can improve or aggravate coking in the heater, so they should be well planned to avoid unpleasant surprises.

Most of the times the original heater structure is not modified during revamp but several parts of the heater can be modified or new parts added in order to achieve the revamp objectives. We will discuss several possible alterations to an existing heater and its' consequences in terms of coking.

Furthermore, two techniques for remediation of coking problems are presented, the use of tube inserts and heat shields.

5.1. Tube inserts

This section includes information that is part of TOTAL know-how, therefore due to its confidential nature it will be presented in a separate volume, which will be available only to TOTAL and the members of the jury that will evaluate this work.

5.2. Heat shields

When, as a result of increased duty or poor burning conditions, sections of tube in the radiant section are overheated and hot-spots are observed, ceramic heat shields wrapped on the tubes can be successfully used to lower the tube's temperature with no reduction in heater performances.

Heat shields act as an additional resistance to heat transfer, they have the same effect as external tube fouling but the thermal resistance can be adjusted. The material used as heat shield is a ceramic fiber fabric layer that can be made in different thicknesses to achieve the desired thermal resistance. There are several different ceramic fibers on the market that can be used in this application. They have low thermal conductivities in the order of 0.1 to 0.2 W/m.K, typical fabric thicknesses are 1, 2 and 3 mm.

Before applying the ceramic fabric, tubes must be clean by the methods described in section 3.6 or at least brushed and dusted. There are two methods of application depending on the system chosen.

The ceramic fiber fabric layer can be glued with a silicate based adhesive and fixed in place with clamps and 2 mm Inconel wire. The overall shield with clamps and wire is coated for further protection. **Carburandum** supplies heat shields with this kind of fixation method.

Tie cords of the same material as the shields or ceramic buttons can be used to fix the shields in place. This allows for easy removal of the heat shields since no glue is used. **Insulcon** provides heat shields with this kind of fixation methods.

Heat shields can be mounted covering the whole tube circumference, 240º, 180º or even less, depending on the heat flux distribution. Only the area affected by overheating and hot-spots needs to be protected. This permits using the rest of the surface for maximum heat transfer. Heat shields usually only cover from 1/3 to 2/3 of the tube circumference and the heat flux is homogenized around the tube.

After heat shield installation there is no effect on heater performance, even though up to 5% of the coil surface in the radiant section can be protected by the ceramic shields. No bridge wall temperature increase has been observed after heat shield installation.

5.3. Air pre-heater

Installing an air pre-heating system is the most popular heater revamp and one of the easiest ways to improve capacity [51]. An air pre-heater can increase radiant section heat duty 10% to 25%, while each 20 ℃ drop in stack tem perature roughly translates into 1% increase in heater efficiency, meaning a total boost in thermal efficiency of 8% to 18%.

Usually stack temperature is in the range of 340 ℃ to 425 ℃ in vacuum heaters. Since it can be brought down to the acid mist condensation temperature, the potential for savings is very big. This is a very attractive investment if the heater size is 10 MW or more and the stack temperature is above 340 ℃.

The installation of an air pre-heater is a major revamp since it also entails the installation of forced draft burners, which involves new burners, blowers and/or fans and ducts for flue gas and combustion air. Besides the costs, enough space to mount all the new equipment is needed, especially under the heater, for forced draft burners and the combustion air duct system. An air pre-heating system is not installed in all heaters because the investment cost of both the pre-heater and the forced draft air system is not always justifiable.

Types of air pre-heaters are presented and discussed in section 2.10.2. But **the regenerative type, although larger and costlier is simpler, requires less maintenance, resists corrosion and needs no power**.

The heat recovered from the flue gas is used to heat the combustion gas before it is fed to the burners. This increases the firing rate and therefore the heat duty in the firebox. This must be taken into account when sizing the burners and the firebox. The increased firing rate will raise radiant heat flux and tube temperatures, so the heater should be re-rated for the new conditions to check for constraints. If attention is not paid to this problem, the heater can develop serious coking problems after revamp.

5.4. Convert a natural draft heater to forced draft

This type of revamp is always done when installing an air pre-heater. It can be sometimes done alone but it is unusual, since it will not provide an economic return on the investment by itself as it would with an air pre-heater. The advantage of installing a forced draft system is in terms of better heater control and flame stability, therefore improving operation.

Natural draft burners require more excess air (see table 3.1) and have longer flames than forced draft burners because of worse air/fuel mixing. Mixing is promoted by air velocity, which is induced by the available air pressure at the burner’s inlet. Since air/fuel mixing improves with pressure drop, there is a limit to the level of mixing we can achieve with natural draft burners.

Forced draft burners provide:

- More turbulence in the firebox, resulting in more uniform tube heating
- More efficient combustion, providing fuel savings
- Possibility to increase the firing duty
- Smaller more stable flames, resulting in less probability of flame impingement

Installing forced draft burners will help reduce and stabilize the flames avoiding flame impingement. The use of a closed combustion air system will also allow improved control of excess air, which will reduce fuel consumption. In some heaters plagued by flame impingement, insufficient draft or excess air control problems, this can be a justifiable solution even without the installation of an air pre-heater.

There is however a possible downside, forced draft burners lower the flame height and therefore the position of the longitudinal peak heat flux. It should be verified if the shift of the peak heat flux won’t increase the coking rate.

When considering changing burners have this in mind: since burners form the heart of a furnace, the process of implementing new ones should always be tried cautiously.

The conversion of a natural draft heater to forced draft is a major revamp since it entails the installation of new burners, blowers and/or fans, a closed circuit for combustion air circulation and a control system requiring a minimum of instrumentation. Besides the costs, enough space to mount the new burners and the combustion air closed circuit is needed, especially under the heater.

5.5. NOx reduction

This section includes information that is part of TOTAL know-how, therefore due to its confidential nature it will be presented in a separate volume which will be available only to TOTAL and the members of the jury that will evaluate this work.

5.6. Convection section

All recent fired heaters are built with a convection section, but this is not always the case in old heaters. If a heater doesn’t have a convection section, one can be added. Since flue gas temperature can be brought down to within 50 ºC of the oil inlet temperature the potential for savings is very big. This is an attractive investment if the heater size is 15 MW or more in size and the oil inlet temperature is low.

If a convection section already exists there are still several revamping options. One of the options is adding more tubes. The convection section height is normally designed to accommodate two additional tube rows. If more tubes rows are required than the old convection section can hold, the section can be lengthened, provided the heater structure can support the additional charge.

If flue gas temperature is still high after the convection section and there is use for low pressure steam, a steam generator or super-heater can be installed on top of the convection section. If steam coils are already in place, another option is to convert their tubes to oil service if the steam they produce can be generated elsewhere. This can be used to increase heater capacity.

The heat transfer area can be increased adding extended surface to the tubes. This can be done with studded tubes for fuel oil fired heaters or finned tubes in fuel gas fired heaters. These tubes exchange respectively 2 to 3 times and 8 to 11 times more heat than bare tubes.

Increasing the convection section heat transfer has several advantages. It can increase heater efficiency and also by transferring more heat by convection, lower the radiant section heat duty, which translates into lower heat fluxes and tube temperatures. This can de-bottleneck a heater with high radiant section heat flux or allow for increased capacity.

5.7. Alter tube coils

Heater operation may be changed to increase vacuum unit production or yield. This involves increasing one or more of the following: feed rate, firing rate, outlet temperature and coil steam. These changes will alter flow regimes, temperatures, and vaporization profile inside the heater's tubes, which will have consequences such as increased pressure drop and higher coking rates.

When pressure drop and/or coking are too high, in order to return the heater to acceptable operation, changes in tube size and size transitions can be made. In addition the number of passes and coil steam injection can be altered. The use of tube inserts or heat shields in some tubes may also help. The objective of these changes is to minimize coking with acceptable pressure drop, as discussed in sections 2.3, 2.6, 5.1 and 5.2.

5.8. Conclusions

Sometimes, as a result of increased duty or poor burning conditions, sections of tube in the radiant section are overheated and hot-spots are observed. Ceramic heat shields wrapped on the tubes can be successfully used to lower the tube's temperature with no reduction in heater performances.

Installing an air pre-heating system is the most popular heater revamp and one of the easiest ways to improve capacity. The regenerative type air pre-heater type, although larger and costlier than others is simpler, requires less maintenance, resists corrosion and needs no power.

When considering changing burners, either because of NOx limits or to install forced draft, have this in mind: since burners form the heart of a furnace, the process of implementing new ones should always be tried cautiously.

6. Troubleshooting

While at DAT, I was given the task of simulating two fired heaters plagued by severe coking problems using simulation software FRNC-5 in order to try to troubleshoot the heaters. The heaters were the Leuna VDU unit heater and the Donges visbreaker unit heater. The history of both heaters, analyses of the problems and their likely causes as well as possible solutions are presented in this chapter.

6.1. Leuna VDU heater

6.1.1. Heater history

The 611-H1002 vacuum heater at Leuna is a 6 pass vertical box heater with 32 radiant section tubes in each pass and a large convection section. The radiant coil is constituted by 24 single-fired tubes connected by an internal crossover to 8 double-fired tubes. The furnace heats 645 ton/h of very heavy atmospheric residue, fed directly from the atmospheric distillation unit, from 344 ºC to 414 ºC. FRNC-5 simulation results at the heater normal operation conditions can be found on figures A.5 and A.6 in the annex section.

The heater has been in operation since 1997 and has suffered from shorter than expected run lengths. This is due to high skin temperatures caused by coking at various locations in the radiant coils, especially on the top part of tubes but also in the inlet and outlet tubes of each pass. Initial efforts were made to limit the high coking rates, including the removal of certain tube guides, increased coil steam flow and additional steam injection points without significant results.

In 2005 additional modifications were made, the burners were modified with a redistribution of gas injection in an attempt to change the vertical heat flux profile and steam ejectors were installed inside the firebox with the goal of increasing the recirculation of flue gas to locally decrease skin temperatures. A study of the impact of these measures was made by CReS [59]. The modifications made to the burners (addition of a central tip for fuel gas injection) do not seem to have a great impact on the thermal flux. Even if it seems to slightly reduce the peak heat flux at 8.5m, it has no impact at 11m where the coke problem exists.

By directing steam ejectors using an infra-red camera to visualize the hot and cold zones generated by the gas flow on the tubes, the refinery succeeded in reducing the temperature of the points with highest skin temperatures out of alarm and increase the duty of the furnace. However, this is a remediation measure, not a solution. Furthermore, the use of steam in the firebox comes at a cost, even though there is a 1.5 MW fuel economy because of the improved heat transfer the steam must be produced, using 2.95 MW and heated in the firebox up to the flue gas temperature, consuming another 1.8 MW.

In late 2005 a study was commissioned to Shell Global Solutions International B. V. [9]. They found that heat fluxes higher than 100 kW/m^2 were experienced by the top of the return bends due to a high heat flux re-radiated by the unshielded refractory roof of the furnace. This caused film temperatures in the top tube bends to rise above 470 ºC, so coke formation was predicted to be severe. The high skin temperatures in the outlet tubes were likely caused by tube thickness, which is twice that of all the other tubes.

After analyzing furnace operating data, Shell Global Solutions modeled the heater using a proprietary software and recommended that coil steam be injected at the radiant section inlet instead of at the inlet of tube 24, at a rate of 0.3% wt of the feed rate (the previous rate was 0.7% wt – 0.8% wt) to improve run length. They also presented three options to decrease peak heat fluxes:

- Install heat shields on the top return bends and internal crossover tubes
- Increase the length of the radiant section tubes such that the top return bends are closer to the refractory roof
- Utilize a section of the steam coils in the convection section to pre-heat the feed so that it enters the radiant section at a higher temperature.

The refinery changed the steam coil injection point and flow rate according to Shell Global Solutions' recommendations and installed heat shields on the top return bends. Nevertheless, high skin temperature and short run length continued, averaging 2 years. Although the top return bends are now protected, the top parts of the tubes still present high skin temperatures.

6.1.2.Problem analysis

This heater has very stable operation due to the integration between the atmospheric and vacuum units. The crude processed in the refinery is Russian crude received directly by pipeline, therefore with very small changes in composition. Firing rate is also stable because the heater is fired by natural gas instead of refinery gas. This eliminates most of the usual causes of coking due to operational upsets. The next step was to investigate possible design problems.

Flow is equally distributed to all passes but passes have different outlet temperatures raging from 412 ºC to 418 ºC. This indicates that heat flux imbalances exist and that the passes don't all receive the same amount of heat. This could be the cause of the problem but all passes have coking problems, although two of them are more severely affected, the ones with high outlet temperature.

The heater was simulated with FRNC-5 fired heater simulator to evaluate conditions inside the tubes and also to test different scenarios. FRNC-5 simulations showed a mass velocity of 1600 kg/m^2.s, average heat flux of 26 kW/m^2 in single-fired tubes and 37 kW/m^2 in double-fired tubes, all film temperatures below 440 ºC and nucleate boiling regime throughout the entire radiant section. All these parameters are excellent, so simulation results don't show any heater problems.

The next study was made using FRNC-5 simulations to see the influence of changes to the heater operation to try and find a better operational point for the heater. I studied the influences of feed rate, coil steam rate and outlet temperature. The conclusions from this study are that feed rate has little influence on the coking rate and coil steam was found to be already optimized at 0.3% wt of the feed rate. Decreasing the heater outlet temperature can reduce temperatures in the heater, including film, and therefore reduce the coking rate. The problem with this solution is lost revenue from reduced distillate production in the VDU column and the refinery already uses this option when a run length extension is necessary.

The coking problem is located in the upper part of the tubes with a maximum at the 11m level, which is approximately the tube length. Heat flux measurements by CReS did not show very elevated heat flux, the values reported by them [59] are within the usual. The question is did their measurements account for re-radiation from the refractory roof?

To measure heat fluxes they used a total heat flow meter and an ellipsoidal radiometer. The first measures the total absorbed heat flux and the second only measures the incident radiant heat flux. So the total heat flow meter would be the best equipment to use to evaluate heat flux in the top part of the tubes. Due to the fact that CReS only had a straight total heat flow meter probe, measurements had to be made where view ports were available and the highest level in this furnace with view ports was 8.5 m. A narrow angle radiometer would also have been adequate but CReS did not have that type of equipment.

So heat flux measurements above 8.5 m, including at 11 m where the problem is most marked, were made with an ellipsoidal radiometer. This type of equipment is best used to measure directly incident radiation, so the measures taken with it were essentially of flue gas radiation. Radiation received at sharp angles, such as from the roof above the tubes, don't have a big influence on the measured radiant heat flux.

Therefore it is possible that a significant radiant heat flux from the refractory roof was not accounted for in the heat flux measurements. If there is indeed a significant radiant heat flux from the refractory roof, it is also possible that it is responsible for the coking problem in the top part of the tubes. The heat shields installed only protect the top half of the tube return bends.

Another possible cause, although less likely, is two-phase flow regime change in the down-flow tubes. In some situations phase separation on the top return bends occur and an unstable flow develops in the top part of down-flow tubes, increasing film temperatures. This is known to occur in long tubes due to a significant difference in hydrostatic pressure and is favored by low fluid velocity. Since tubes in this heater are only 11 m long and fluid velocity is high it is unlikely that this is the cause.

6.1.3. Likely cause

The most likely cause is high heat flux due to significant re-radiation from the refractory roof, with two-phase flow regime change in the down-flow tubes top part as a less likely cause. This can be easily verified by heat flux measurement if adequate equipment is provided to CReS or if an outside company is contracted.

An option to discard the two-phase flow regime change is to compare heater diagrams and thermograph survey data to see if the tubes with high skin temperatures are all down-flowing tubes.

6.1.4. Possible solution

If the problem is confirmed to be high heat flux due to significant re-radiation from the refractory roof, a possible solution is to cover the refractory roof with tubes. The total heat transfer area in the radiant section can be increased with these new tubes or maintained by changing the tube center to center spacing. The first option will increase the heater's capacity and the second will reduce radial heat flux imbalance, lowering radial peak heat fluxes to the tubes.

If the problem is two-phase flow regime change in the down-flow tubes it may be possible to change this by altering the feed flow or the coil steam injection but it doesn't always work. Several coil steam injection rates and locations have already been attempted in this furnace without any visible results.

6.2. Donges Visbreaker heater

6.2.1. Heater history

The 691-F1001 visbreaker heater at Donges is a 4 pass vertical box heater with 56 radiant section tubes in each pass and a convection section. The radiant coil is constituted by 19 double-fired tubes connected by an external crossover to an additional 37 single-fired tubes. The furnace heats between 130 and 200 ton/h of vacuum residue that is fed to the heater from a tank at 310 ºC and heated up to 460 ºC. FRNC-5 simulation results at different heater operation conditions can be found on figures A.7, A.8 and A.9 in the annex section.

The heater was commissioned on June 1986 and had operated without significant convection section problems until the 14th of July 2006, when the rupture of a first row shock tube forced an emergency stop. The ruptured tube was the last shock tube of pass four and the correspondent tube in pass two also showed an important fissure that could also have soon led to a tube rupture.

The tube rupture followed an incident two days before, when the heater suffered an emergency stop due to a feed pump malfunction without secondary pump start. During this incident the feed was cut and the oil stayed in the tubes some five minutes at high heat flux, which no doubt led to intense coking inside the tubes and subjected the tubes to high skin temperatures close to the flue gas temperature.

The affected tubes were replaced and the ruptured sections sent to CReG to the corrosion/materials lab for analysis by a metallurgic expert. A report was issued [60] and the essential of that report is summed here. Both tubes showed and important layer of coke with thicknesses between 13 and 17 mm. Two types of coke of different composition and structure were observed, a layer of friable coke with 1 to 7 mm covering only the lower wall side exposed to the flue gas, covered by a 5 to 10 mm layer of compact coke covering the entire circumference of the wall that may have formed due to the incident two days before the accident. This cannot be confirmed and the coke may have formed due to another problem.

Figure 6.1 shows the ruptured section of shock tube and the coke and metal thicknesses around the tube's circumference in a zone near the fissure.

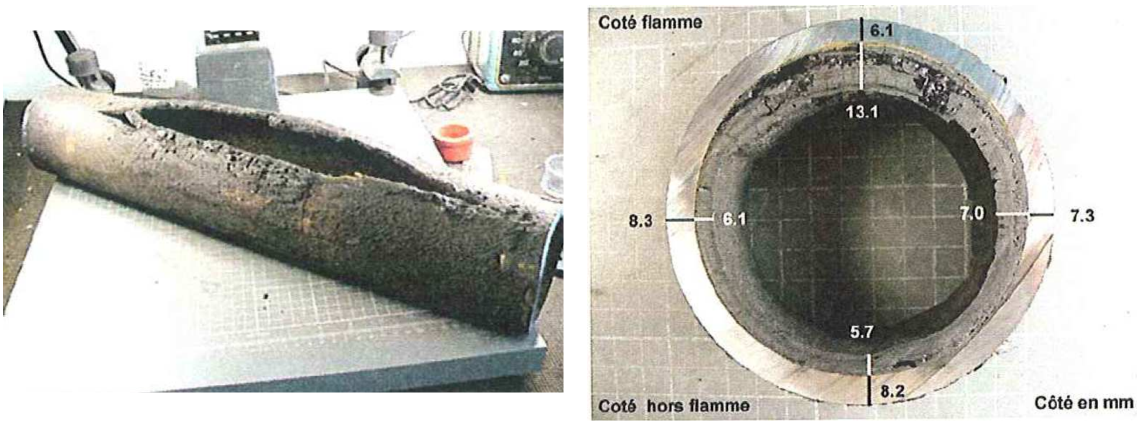


Figure 6.1 - Ruptured shock tube section and coke layer near the fissure [60]

What can be stated as a fact is that the metal degradation process, including bulging and thinning of the tube walls on the lower side of the tube, was already advanced before the incident. Creep deformation and metal carburization were also observed by microstructure analysis but could be caused by the accident. In October 2005, time of the last decoking stop, the tubes already showed a subsidence between tube supports.

To CReG, a metal degradation like this is surprising in the convection section, so they proposed the installation of skin thermocouples to follow up shock tube skin temperatures in order to better understand the phenomenon. The skin temperature thermocouples recommended were installed and the heater decoked with air/steam. The new skin temperature thermocouples showed rapidly increasing shock tube metal temperatures in all four passes. This has limited the heater run length, which has decreased considerably from the runs before the accident. The heater has also latter been decoked by pigging without significant change in the rate of skin temperature rise.

The conclusion is the accident cannot be explained only by the incident two days before. It may have been the final straw that precipitated the tube rupture but it was not the only cause. Two pigging reports, one made before and another one after the accident indicate two types of coke, one hard and one soft, the same type of coke layer found in the analysis of the ruptured tube sections. Coke thicknesses were significant, not as large as the ones measured in the ruptured tubes but even so 7 mm coke thicknesses were registered.

There is another indication that the problem already existed before the accident. Shock tube skin thermocouples were only installed after the accident but radiation inlet skin thermocouples were installed at the time and these temperatures have the same tendency as the shock tube skin temperatures. Historically, radiation inlet skin temperatures have been high, at times even higher than nowadays, so it is likely that the shock tubes were already affected by high skin temperatures and the situation went unnoticed. The fact that one of the tubes showed a slight deformation before the accident is indicative of high skin temperatures. The additional coke deposited during the feed cut incident precipitated the rupture but I believe it wasn't the only cause.

6.2.2.Problem analysis

Heurtey Petrochem was commissioned to study the problem and a report complete with several recommendations was made [61], pointing out several situations that can favour coking in the shock tubes:

- high process severity
- severe convection section fouling by mineral deposits
- possible significant excess air variation
- repeated charge rate and outlet temperature variations
- presence of salts in the feed
- possibility of formation of an oil/water emulsion

These possible causes as well as some others were investigated by analyzing refinery operation data and by performing FRNC-5 simulations under different scenarios. The first step was to analyze refinery operation data available through the PI Process Book interface, which allows direct access to the DCS sensor and controller historic data. Available skin temperatures and all other measured variables available that could help explain coking rate increases were analyzed, including: viscosity; heater inlet, radiation inlet and outlet temperatures; condensate injection rate; charge feed rate; excess air; fuel rate and bridgewall temperature. A graph showing the evolution of shock tube skin temperature and of some other variables is presented in figure 6.2. The red horizontal line marks 470 ºC.

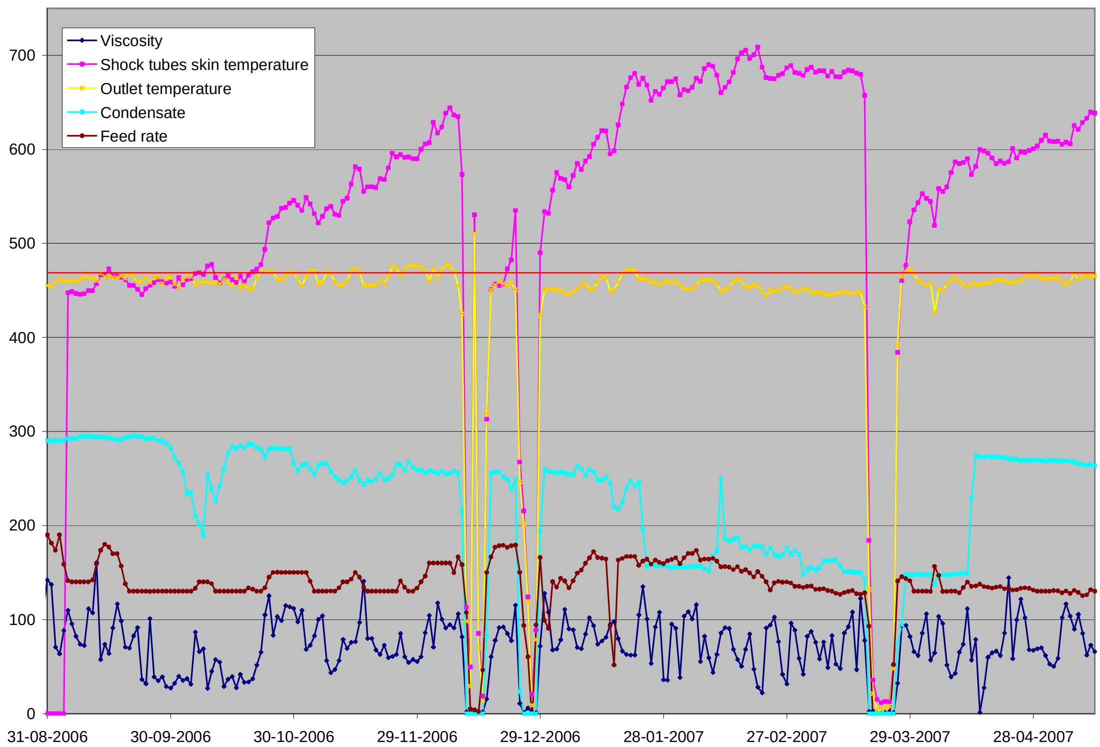


Figure 6.2 - Heater daily averages after the installation of shock tubes skin thermocouples

From refinery data analysis it is clear that the average outlet temperature has risen by about 10 ℃ from the middle of 2005. During the last year and a half the heater has been operating at higher severity. Feed rate also changes considerably, there are typically three different feed rates: 130 ton/h, 160 ton/h and 200 ton/h.

The simulations with FRNC-5 allowed exploring the effect of different problems and operational variables in the shock tube coking rate. This in conjunction with refinery data permitted to prove or discard their influence in coking. I will now briefly present the studies made with FRNC-5:

Condensate feed rate

Simulations of the visbreaker heater, varying the condensate feed rate at different charge feed rates, show that it can change the boiling regime at low charge rates with significant influence in skin temperatures. The heat transfer coefficient more than doubles, lowering skin temperatures as much as 25 ºC. This effect can explain the 15 ºC skin temperature drop verified when a condensate feed cut test was made. I consider a boiling regime change a more likely explanation than the formation of an oil/water emulsion as advanced by Heurtey.

Oil feed rate

When oil feed rate is increased, mass velocity will increase with a beneficial effect in heat transfer but the added amount of oil to heat increases the furnace heat duty and therefore the heat fluxes. The overall effect is a slight increase in coking rates but it is not enough to explain the very high skin temperature rise in the shock tubes. The fact that charge rate varies frequently as well as significantly can contribute to increase coke deposits due to temporarily high heat flux and two-phase flow regime change. But once more it is not enough to explain the very high skin temperature rise in the shock tubes.

Heater outlet temperature

As all temperatures in the furnace are linked to heater outlet temperature, if it is increased it will affect film as well as skin temperatures. Operating above 460 ºC is considered high severity with expected high coking rates. As can be seen in figure 6.2, outlet temperature has been frequently kept above 460 ºC since august 2005, sometimes even passing 470 ºC. By itself it is not reason enough to explain the problem but it will nevertheless help coking.

Convection section fouling

The convection section is severely fouled in its upper part, on the steam superheater tubes, by a mineral deposit. Only the central part, directly under the stack is fouled, this forces the flue gas around it and increases convection in the non fouled part of the tubes. Flue gas velocity measures by CReS show that some flue gas still passes through the center of the convection section but with low velocity compared to the flue gas flow in the rest of the section (6 m/s vs 1.5 m/s). The measures made by CReS were made directly below the fouled zone, since the shock tubes are more than 1.5 m below they won't be as affected.

Simulations with FRNC-5 show that even if only half the convection section heat transfer area is effective there is no effect on the coking rate and skin temperature would rise only by 7 ºC due to the added heat flux.

Charge quality

Charge quality changes frequently but it was not thought to be significant since the vacuum residue that feeds the visbreaker is stored in a tank before being fed to the unit.

Based on refinery data analysis, DAT recommended to the refinery the return to old operational values of lower severity and high feed rate. A new run was started on the 10th of July with an outlet temperature of 460 ºC, a feed rate of 150 ton/h, later increased to 160 ton/h and high excess air to lower radiation heat flux to the shock tubes. Despite keeping a stable run with conditions thought to be the best to limit coke in this heater, on the 14th shock tube skin temperatures started increasing rapidly as usual.

After carefully looking at the heater data during those first four days, an abrupt change in feed viscosity was noticed. Viscosity changed from 87 cP to 262 cP in less than an hour indicating that there was a change in charge quality, which became much heavier and therefore with a much higher coking potential. The heavier oil will also vaporize at higher temperatures, increasing film temperatures, lowering fluid velocity and changing two-phase flow regimes.

During the start of a bitumen production run the vacuum residue quality must be regulated and the residue produced out of specification is fed to the visbreaker feed tank. These runs are frequent and they produce very heavy vacuum residues. Vacuum residue is analyzed on-line by a viscosimeter at the column outlet and another measure the heater feed at the inlet. When we look at both viscosity values, it is evident that the visbreaker feed sometimes follows the vacuum column residue's viscosity. This indicates that the visbreaker feed tank is not adequately agitated and stratification occurs. This causes a "poured drain" situation where there is almost no homogenization in the tank and the visbreaker will be affected by abrupt changes in charge quality.

High heat flux

This heater has an unusual configuration, each cell has its own flue gas opening and the flue gas is channelled to the convection section through large 90º curve ducts lined with refractory. An infrared thermograph survey shows that the high shock tube skin temperature problem is located in the zones directly in front of the flue gas ducts [62].

Flue gas is hot enough in the ducts to still radiate a significant amount of heat to the shock tubes but this is normal in all heaters. The problem is the refractory covering the flue gas ducts, refractory has an emissivity of 0.85 or more, while flue gas emissivity is less than 0.40. Shock tubes in this heater receive a much higher radiant heat flux than usual, which results in high skin and film temperatures and accelerated coking.

Excess air

Since February 2007 the refinery has been using high excess air, more than 50%, which they report, helps to lower shock tube skin temperatures. Normally the opposite would be expected, more excess air lowers the bridgewall temperature and increases the volume of air, all of which reduces radiant heat transfer and increases convection. It would be expected then that heat flux would increase in the shock tubes. Nevertheless, shock tube heat flux is reduced, as confirmed by lower skin temperatures, strengthening the theory about high radiant heat flux to the shock tubes. Lower flue gas temperature lowers the radiant heat flux from the flue gas ducts refractory to the shock tubes.

6.2.3. Likely causes

The most likely cause is very high heat flux to the shock tubes from the flue gas ducts refractory. Their cherry colour indicates they are very hot. This problem is no doubt helped by the high outlet temperature and frequent changes in the feed rate. Increased pressure drop due to coke will also raise the oil boiling temperature which has the same effect in film temperatures resulting in additional coking.

Abrupt changes in charge quality during vacuum unit bitumen run starts seem to initiate severe coking in the shock tubes. The problem may be high asphaltene content or just changes in vaporization profiles and two-phase flow regimes.

6.2.4. Possible solution

Heat shields can reduce heat flux to the shock tubes. The bottom row should be completely covered by heat shields and the bottom half of the next row should also be protected. This is a simple and cheap measure that should alleviate the skin temperatures and coking. Heat shields are easy to install and if necessary they are easy to uninstall. The only inconvenient is the need to install scaffolding inside the heater in order to reach the shock tubes.

This measure was discussed in a meeting at Donges with refinery personnel and other intervenients on the heater troubleshooting. It was accepted as the best solution after review of troubleshooting data and proposed to the refinery for application during the next stop.

Mixers in the visbreaker feed tank may help homogenizing the feed to the heater to avoid abrupt changes in charge quality. Another solution may be the use of a separate tank to store very heavy vacuum residue from vacuum unit run starts. This heavy residue would be slowly added to the visbreaker feed in order not to change the charge too much.

7. Recommendations

Coking in fired heaters is a complex matter and this work cannot cover all its relevant aspects. Therefore I recommend further studies in this area, specifically on four subjects.

An economical analysis of the different design specifications would enable us to find the cheapest fired heater design. The same results in limiting coke can be achieved by increasing the mass velocity or decreasing the heat flux but the cost relation of implementing these measures is not known. By analyzing the true cost of altering mass and heat flux, if possible working with a fired heater engineering company like Heurtey Petrochem, we could obtain not only non-coking design specifications, but also the assurance that those would lead to the cheapest heater design.

Several charge quality parameters have been linked to coke formation, like asphaltene, sodium, sulphur and metal content and peptization state, oxidation level, acidity and asphaltene solubility or flocculation tendency. Since there are several vacuum heaters in the group with important charge variation, one of them might be followed with extended charge analysis to verify which of the charge parameters suggested have more influence in coking, as measured by the increase in skin temperature after influences such as heat duty and outlet temperature variation have been deducted.

Study the “dry” VDU heaters (without the use of injected steam) in the group to compare their mass and heat flux, as well as maximum film temperatures, with “wet” heaters. The current study did not evolve in that direction, but this would allow us to know the real effect of coil steam and also to better evaluate other design parameters that might be different in that type of heater.

The influence of mass velocity in residence time and fluid velocity and therefore on coking rates is easily calculated. But the reduction effect of wall shear stress on coking rates because of increased fluid velocity is not yet quantifiable. Such a study could be made in a simple “Fouling tester” apparatus such as the one described by Mr. Toussaint [21]. Such an apparatus could also be used in the study of charge quality influence described above, since in actual heater operation too many variables change and therefore results are not always reliable.

As a study of the nature of this work is important to avoid coking problems, it would perhaps be appropriate to do so for other fired heaters in residue service like visbreaker unit heaters.

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9. Annex

Maximum Heat Release per Burner		Minimum Clearance: A Vertical to Centerline Roof Tubes or Refractory (Vertical Firing Only)		B Horizontal to Centerline Wall Tubes from Burner Centerline		C Horizontal from Centerline of Burner to Unshielded Refractory		D Between Opposing Burner (Horizontal Firing)	
MW	million Btu/hr	meters	feet	meters	feet	meters	feet	meters	feet
				Oil Firing					
1.2	4	3.7	12	0.8	2'9"	0.6	2'0"	4.9	16
1.8	6	4.9	16	1.0	3'3"	0.8	2'6"	6.7	22
2.3	8	6.1	20	1.1	3'9"	0.9	3'0"	8.5	28
2.9	10	7.3	24	1.3	4'3"	1.1	3'6"	9.8	32
3.5	12	8.5	28	1.4	4'9"	1.2	4'0"	11.0	36
4.1	14	9.8	32	1.6	5'3"	1.4	4'6"	12.2	40
				Gas Firing (Low NO_x burners[c])					
0.6	2	2.7	9'1"	0.6	2'0"	0.5	1'6"	3.9	12
1.2	4	3.9	13'0"	0.8	2'6"	0.6	2'0"	5.9	18
1.8	6	5.2	16'11"	0.9	3'0"	0.8	2'6"	7.9	24
2.3	8	6.4	20'9"	1.1	3'6"	0.9	3'0"	9.8	30
2.9	10	7.5	24'8"	1.2	4'0"	1.1	3'6"	11.8	36
3.5	12	8.7	28'7"	1.4	4'6"	1.2	4'0"	12.8	39
4.1	14	12.9	32'6"	1.5	5'0"	1.4	4'6"	13.8	42
4.7	16	15.5	36'5"	1.8	5"6"	1.5	5'0"	14.8	45
5.3	18	17.2	40'4"	2.0	6'0"	1.8	5'6"	16.4	50

Notes:
[a]For horizontal firing, the distance between the burner centerline and the roof tube centerline or refractory shall be 50 percent greater than the distances in Column B.
[b]For combination liquid and gas burners, the clearances will be based on liquid fuel firing, except when liquid fuel is used for start up only.
[c]For conventional (non-Low NOx) gas burners, a decrease in longitudinal clearance may be allowed. This is achieved by multiplying dimensions in column A by a factor of 0.77 and D by a factor of 0.67.
[d]For intermediate firing rates, the required clearances may be achieved by linear interpolation.

Figure A.1 - Minimum burner clearance for natural draft operation by API Standard 560 [1]

Gas Firing

Maximum Heat Release per Burner, MM Btu/h	UOP Spacing horizontal to centerline wall tube from burner centerline, inches	API Spacing horizontal to centerline wall tube from burner centerline, inches
4	39	30
6	39	36
8	45	42
10	48	48
12	54	54
14	60	60

Oil Firing

Maximum Heat Release per Burner, MM Btu/h	UOP Spacing horizontal to centerline wall tube from burner centerline, inches	API Spacing horizontal to centerline wall tube from burner centerline, inches
4	48	33
6	48	39
8	54	45
10	57	51

Oil Firing- Staged Air Burner

Maximum Heat Release per Burner, MM Btu/h	UOP Spacing horizontal to centerline wall tube from burner centerline, inches	API Spacing horizontal to centerline wall tube from burner centerline, inches
4	54	33
6	54	39
8	60	45
10	63	51

Figure A.2 - Recommended burner clearance by UOP [30]

TABLE 1. RECOMMENDED MINIMUM BURNER-TO-TUBE CLEARANCES		
Burner design heat liberation, million Btu/h	**Clearance, burner-to-tube centerline, ft-in.**	
A. Natural-draft burners:	**Gas**	**Oil**
2	2-6	3-0
4	3-0	3-6
6	3-6	4-0
8	4-0	4-6
10	4-6	5-0
12	5-0	5-6
B. Forced-draft burners:	**Gas**	**Oil**
5	3-0	3-6
10	3-2	3-8
15	3-4	3-10
20	3-6	4-0
25	3-8	4-2
C. High-intensity burners:	**Gas**	**Oil**
15	4-6	5-0
20	4-8	5-2
25	4-10	5-4
30	5-0	5-6
35	5-2	5-8
40	5-6	5-10

Figure A.3 - Minimum burner clearance recommended by Garg and Ghosh [39]

	MAXIMUM HEAT RELEASE PER BURNER (MM BTU/HR)	A VERTICAL TO ℄ ROOF TUBES OR REFRACTORY (VERTICAL FIRING ONLY)	B HORIZONTAL TO ℄ WALL TUBES FROM ℄ BURNER	C HORIZONTAL FROM ℄ OF BURNER TO UNSHIELDED REFRACTORY	D BETWEEN OPPOSING BURNERS (HORIZONTAL FIRING ONLY)
Oil	4	12'	2'-9" (3'-6")	2'-0"	16'
	6	16'	3'-3" (4'-0")	2'-6"	22'
	8	20'	3'-9" (4'-6")	3'-0"	28'
	10	24'	4'-3" (5'-0")	3'-6"	32'
	12	28'	4'-9" (5'-6")	4'-0"	36'
	14	32'	5'-3" (6'-0")	4'-6"	40'
Gas	2	7'	2'-0" (3'-0")	1'-6"	8'
	4	10'	2'-6" (3'-3")	2'-0"	12'
	6	13'	3'-0" (3'-9")	2'-6"	16'
	8	16'	3'-6" (4'-3")	3'-0"	20'
	10	19'	4'-0" (4'-8")	3'-6"	24'
	12	22'	4'-6" (5'-0")	4'-0"	26'
	14	25'	5'-0" (5'-6")	4'-6"	28'

Note 1: For horizontal firing, the distance between the burner centerline and the roof tube centerline or refractory shall be 50 percent greater than the distances in Column B to account for flame uplift due to buoyancy effects.

Note 2: For combination liquid and gas burners the clearances will be based on liquid fuel firing, except when liquid fuel is used for start up only.

Figure A.4 - Minimum clearance for natural draft non-premix burners above API recommendation [63]

Table A.1 - VDU comparative table[5]

		Inlet Temp	Outlet Temp	Vaporization	Excess O_2	Heater efficiency	VGO Cut-point	Flash zone Pressure	Flash zone Temperature	Transfer line temperature drop	Feed rate	Heat duty	Feed specific gravity	T Flue gas	Run length	Reason for unit stop / Observations
		℃	℃	% wt	% vol	%	℃	mbar	℃	℃	ton/h	MW	@15℃	℃	Months	
Donges	DGS	359	405		3	84,3	528-557	67	388	17	520	41	0,951	250-300	60	Maintenance stop
Feyzin	FZN	314	418	64-72	2,9	82,7	555	106/90-115	401/406	12-17	230	28	0,932	314	60	Decoking stop
Grandpuits	GPS	355	415	50-70	3		558	60/47-73	395/380-400	20	257	20		315	35	Maintenance stop
Lindsey	LOR VDU 1	240	355		6	84,7	498	99	350/320-355	5	64	5	1,03	200-250		
Lindsey	LOR VDU 2	300	416		[6]	79,7	564	118/100-125	411	5	461	62	0,936	[6]	30	Decoking stop[6]
Milford Haven	MHR	263	416				602	24/22-26	410	6	236	40	0,923			Dry unit
Port-Arthur	PAR	360	413	54	5	77,8		17	382	31	302	21		376	60	Maintenance stop/Dry unit
Roma	RdR	265	390	37	3	78,6					60	8	0,99	370-430	12	Decoking stop
Flanders	RF	315	420-425	65-70	4,35	82,8	563	60/50-70	404/405-410	15	271	37	0,938	390	24	Expected. Former 1 year
Normandy	RN D2	265	400		4,75	75,8	520	132	386	14	84	12	0,95	465		
Normandy	RN D5	263	410			74,9	577	71	400	10	94	14	0,957			
Normandy	RN D8	276	395				506	145	381	14	112	14	0,968			
Normandy	RN D10	276	395		3-3,5	85,9	515	124	378	17	160	20		237		
Provence	RP	290	415/430	50/75	2		580	50/95	395/410	20	300	45	0,9201/0,9792	215	14	2 different crude type runs
Antwerp	TRA	280	410	67	[7]	77,4	560	60	388	22	508	87	0,943	[7]	60	Maintenance stop[7]
Leuna	TRM	340	416	60	3,5	85,1	553	77	393		637	60	0,959	250-255	24	Decoking stop
Vlissingen	TRN	325	413		3,4	79,8		70			422	43	0,962	365		
Normandy	RN D12	326	415			92	600	105	406	9	667	71	0,937		60	Project phase. Design values
	Average	**300**	**410**	**60**	**3.7**	**81.5**	**553**	**90**	**392**	**9**	**300**	**35**	**0.955**	**320**	**40**	
	Best Value	**360**			**2**	**92**	**602**	**17/50**[8]	**411**	**5**				**200**	**60**	

LOR - VDU2		
	Heater 1	Heater 2
T flue	340	310
Excess O_2	4,8	6,6
T in	300	370
T out	370	416

Antwerp - Pre-heater + 2 Heaters			
	Pre Heater	Heater 1	Heater 2
T flue	420	340	430
Excess O_2	4,9	6,8	5

[5] Most of the data presented in this table comes from the VDU Benchmarking study [32], but there are also data from simulation results and a revamp study [31]

[6] Lindsey's VDU 2 has two heaters in series. See table **LOR – VDU2** for data on the 2 heaters

[7] In Antwerp's VDU there are two heaters and a pre-heat furnace. See table **Antwerp – Pre-heater + 2 heaters** for data on the heaters

[8] The first value refers to "dry" VDUs and the second to "wet" VDUs that use heater coil steam and column bottom stripping steam

In the Leuna VDU heater's FRNC-5 simulation results, coils 21 and 24 represent convection section tubes, coils 11 to 16 represent radiant section tubes and coils 51 to 58 represent the transfer line. Coil 12 is the crossover tube from the last single-fired tube to the first double-fired tube.

COIL SECT ID.	TUBE NO.	BULK TEMP. DEG.C	BULK PRES KPAA	WT. FRAC. VAPOR	BULK VEL. M/SEC	INSIDE H.T. COEFF.	***PEAK TEMPS.*** INNER FILM DEG.C	AVER. WALL DEG.C	OUTER SKIN DEG.C	AVG. FLUX W/M2	PEAK FLUX W/M2	FLOW REGIME	BOILING REGIME	RELATIVE COKING RATE
24	1	346.	592.3	.000	3.0	2093.2	354.	356.	358.	11291.	15310.	TURBULNT	SENSIBLE	.753E+00
24	2	347.	583.1	.000	3.0	2106.3	357.	359.	362.	13959.	18942.	TURBULNT	SENSIBLE	.843E+00
24	3	349.	573.9	.000	3.0	2122.8	361.	364.	367.	17393.	23624.	TURBULNT	SENSIBLE	.971E+00
24	4	351.	564.7	.000	3.0	2143.7	366.	370.	373.	21770.	29606.	TURBULNT	SENSIBLE	.116E+01
24	5	353.	555.5	.000	3.0	2169.5	372.	377.	382.	27522.	37485.	TURBULNT	SENSIBLE	.144E+01
24	6	356.	546.3	.000	3.0	2202.5	380.	386.	392.	35061.	47841.	TURBULNT	SENSIBLE	.191E+01
21	1	358.	537.8	.000	3.0	2177.4	369.	371.	374.	12832.	21603.	TURBULNT	SENSIBLE	.138E+01
21	2	359.	528.5	.000	3.0	2187.9	371.	374.	377.	14021.	23621.	TURBULNT	SENSIBLE	.150E+01
21	3	360.	519.2	.000	3.0	2199.3	373.	377.	380.	15344.	25867.	TURBULNT	SENSIBLE	.164E+01
21	4	362.	509.8	.000	3.0	2215.6	392.	400.	407.	18753.	60518.	TURBULNT	SENSIBLE	.288E+01
21	5	364.	500.5	.000	3.0	2232.9	395.	403.	411.	21815.	63214.	TURBULNT	SENSIBLE	.322E+01
21	6	367.	491.2	.000	3.0	2268.1	399.	408.	416.	33252.	65944.	TURBULNT	SENSIBLE	.374E+01
11	1	369.	545.2	.007	4.8	2291.3	392.	399.	406.	26609.	48817.	ANNULAR	NUCLEATE	.232E+01
11	2	371.	483.9	.007	5.2	2349.0	394.	400.	407.	26568.	48745.	ANNULAR	NUCLEATE	.231E+01
11	3	373.	530.5	.007	4.9	2325.4	395.	402.	409.	26515.	48650.	ANNULAR	NUCLEATE	.260E+01
11	4	374.	470.7	.008	5.4	2386.1	396.	403.	409.	26484.	48595.	ANNULAR	NUCLEATE	.254E+01
11	5	376.	514.6	.007	5.1	2362.9	398.	405.	412.	26425.	48491.	ANNULAR	NUCLEATE	.289E+01
11	6	377.	456.8	.008	5.5	2424.4	399.	406.	412.	26397.	48441.	ANNULAR	NUCLEATE	.281E+01
11	7	379.	497.7	.008	5.2	2400.6	401.	408.	414.	26342.	48342.	ANNULAR	NUCLEATE	.316E+01
11	8	381.	442.0	.009	5.7	2467.1	402.	409.	415.	26305.	48276.	ANNULAR	NUCLEATE	.311E+01
11	9	382.	480.1	.009	5.4	2440.1	404.	410.	417.	26255.	48188.	ANNULAR	NUCLEATE	.346E+01
11	10	384.	426.3	.009	5.9	2510.9	405.	412.	418.	26215.	48116.	ANNULAR	NUCLEATE	.342E+01
11	11	386.	461.3	.009	5.6	2483.9	407.	413.	420.	26164.	48026.	ANNULAR	NUCLEATE	.380E+01
11	12	387.	409.9	.010	6.2	2561.3	408.	414.	421.	26128.	47962.	ANNULAR	NUCLEATE	.370E+01
11	13	389.	441.0	.010	5.8	2533.0	410.	416.	423.	26076.	47869.	ANNULAR	NUCLEATE	.413E+01
11	14	390.	391.0	.011	6.5	2618.7	410.	417.	423.	26043.	47810.	ANNULAR	NUCLEATE	.398E+01
11	15	392.	420.6	.011	6.1	2585.8	412.	419.	425.	25989.	47713.	ANNULAR	NUCLEATE	.446E+01
11	16	394.	371.1	.013	6.9	2683.5	413.	419.	426.	25958.	47659.	ANNULAR	NUCLEATE	.425E+01
11	17	395.	399.0	.012	6.5	2647.3	415.	421.	428.	25903.	47560.	ANNULAR	NUCLEATE	.476E+01
11	18	397.	350.1	.014	7.3	2752.7	416.	422.	428.	25872.	47505.	ANNULAR	NUCLEATE	.452E+01
11	19	398.	376.2	.013	6.9	2717.4	417.	424.	430.	25819.	47411.	ANNULAR	NUCLEATE	.503E+01
11	20	400.	327.8	.015	7.8	2827.9	418.	425.	431.	25785.	47349.	ANNULAR	NUCLEATE	.480E+01
11	21	402.	352.0	.015	7.4	2790.4	420.	426.	433.	25731.	47253.	ANNULAR	NUCLEATE	.534E+01
11	22	403.	304.1	.017	8.5	2926.3	421.	427.	433.	25700.	47198.	ANNULAR	NUCLEATE	.498E+01
11	23	405.	325.7	.016	7.9	2874.6	423.	429.	435.	25642.	47094.	ANNULAR	NUCLEATE	.562E+01
11	24	406.	278.2	.021	9.6	3075.1	422.	429.	435.	25636.	47084.	ANNULAR	NUCLEATE	.485E+01
12	1	406.	267.6	.023	10.1	3137.3	424.	431.	438.	28041.	50920.	BUBBLY	NUCLEATE	.493E+01
13	1	408.	288.2	.022	9.5	3091.8	425.	431.	437.	36286.	46126.	ANNULAR	NUCLEATE	.542E+01
13	2	410.	241.0	.028	11.8	3330.9	425.	431.	438.	36255.	46085.	ANNULAR	NUCLEATE	.477E+01
13	3	412.	251.8	.028	11.3	3306.8	428.	434.	440.	36157.	45955.	ANNULAR	NUCLEATE	.532E+01
13	4	414.	204.1	.036	14.6	3603.1	428.	435.	441.	36111.	45894.	ANNULAR	NUCLEATE	.466E+01
13	5	417.	206.1	.037	14.7	3623.9	430.	437.	443.	36022.	45778.	ANNULAR	NUCLEATE	.501E+01
14	1	417.	152.2	.064	25.7	4388.6	428.	435.	441.	37835.	45808.	ANNULAR	NUCLEATE	.329E+01
15	1	418.	110.9	.099	24.8	3211.0	433.	441.	448.	37338.	45476.	ANNULAR	NUCLEATE	.493E+01
16	1	416.	74.2	.168	36.8	3018.3	433.	447.	460.	35602.	44973.	ANNULAR	NUCLEATE	.494E+01
51	1	413.	51.5	.224	22.6	.0	413.	413.	413.	0.	0.	ANNULAR	SENSIBLE	.000E+00
52	1	411.	42.1	.258	30.9	.0	411.	411.	411.	0.	0.	ANNULAR	SENSIBLE	.000E+00
53	1	409.	38.3	.274	35.5	.0	409.	409.	409.	0.	0.	ANN.MIST	SENSIBLE	.000E+00
55	1	408.	33.5	.296	36.5	.0	408.	408.	408.	0.	0.	ANN.MIST	SENSIBLE	.000E+00
56	1	404.	22.1	.363	34.2	.0	404.	404.	404.	0.	0.	ANNULAR	SENSIBLE	.000E+00
57	1	403.	19.9	.380	39.3	.0	403.	403.	403.	0.	0.	ANNULAR	SENSIBLE	.000E+00
58	1	399.	13.8	.434	62.7	.0	399.	399.	399.	0.	0.	ANN.MIST	SENSIBLE	.000E+00

Figure A.5 - Leuna simulation results - Summary of tubeside information[9]

TERMINAL CONDITIONS OF EACH COIL SECTIONS***

COIL SECT ID.	POINT INFO. AT	PROC. FLOW SEQ.	PRESSURE (KPAA)	TEMPERATURE (C)	ENTHALPY (J/G)	WEIGHT FRACT. VAPOR	VELOCITY (M/S)	CRITICAL VELOCITY (M/S)	MASS VELOCITY (KG/SEC/M2)	HOMOGEN. REYNOLDS NUMBER	FLOW REGIME	BOILING REGIME
24	INLET	1	600.0	344.8	728.49	.000	3.0	.0	2313.348	497900.	TURBULNT	SENSIBLE
24	EXIT	1	545.6	356.6	762.66	.000	3.0	.0	2313.348	548477.	TURBULNT	SENSIBLE
21	EXIT	2	490.5	367.4	793.73	.000	3.0	.0	2313.348	596426.	TURBULNT	SENSIBLE
11	EXIT	3	271.2	405.5	917.21	.022	9.9	47.0	1606.247	2501282.	ANNULAR	NUCLEATE
12	EXIT	4	262.4	406.1	919.44	.023	10.4	47.0	1606.247	2625494.	BUBBLY	NUCLEATE
13	EXIT	5	195.3	416.6	953.14	.040	16.0	47.9	1606.247	3743649.	ANNULAR	NUCLEATE
14	EXIT	6	103.2	414.9	960.23	.102	47.4	56.8	1606.247	5627204.	ANNULAR	NUCLEATE
15	EXIT	7	95.3	416.3	969.39	.122	33.4	60.2	921.771	5938155.	ANNULAR	NUCLEATE
16	EXIT	8	54.0	413.6	979.89	.215	58.6	70.8	646.130	7072859.	ANNULAR	NUCLEATE
51	EXIT	9	45.0	411.3	979.89	.247	27.9	74.5	231.557	5050804.	ANNULAR	SENSIBLE
52	EXIT	10	38.7	409.6	979.89	.272	35.0	77.2	231.557	5426630.	ANNULAR	SENSIBLE
53	EXIT	11	34.4	408.3	979.89	.292	41.4	79.1	231.557	5578757.	ANN.MIST	SENSIBLE
55	EXIT	12	22.2	403.9	979.89	.363	63.8	85.4	196.909	9218607.	ANN.MIST	SENSIBLE
56	EXIT	13	20.6	403.2	979.89	.374	37.6	86.4	105.160	10299550.	ANNULAR	SENSIBLE
57	EXIT	14	14.1	399.4	979.89	.430	60.8	91.0	105.160	10447270.	ANNULAR	SENSIBLE
58	EXIT	15	10.0	396.0	979.89	.478	93.8	95.0	105.160	10842010.	ANN.MIST	SENSIBLE

Figure A.6 - Leuna simulation results - Terminal conditions of each coil section

[9] Information applies to average conditions in exit increment of each tube. Tubes are numbered from coil inlet

In the Donges visbreaker heater's FRNC-5 simulation results, coils 21 to 23 represent convection section tubes with extended surface, coil 24 represents the shock tubes and coils 11 and 13 represent radiant section tubes. Coil 12 is the external crossover tube from the last double-fired tube to the first single-fired tube.

COIL SECT ID.	POINT INFO. AT	PROC. FLOW SEQ.	PRESSURE (KPAA)	TEMPERATURE (C)	ENTHALPY (J/G)	WEIGHT FRACT. VAPOR	VELOCITY (M/S)	CRITICAL VELOCITY (M/S)	MASS VELOCITY (KG/SEC/M2)	HOMOGEN. REYNOLDS NUMBER	FLOW REGIME	BOILING REGIME
21	INLET	1	1759.1	309.1	643.43	.000	1.9	.0	1574.567	112040.	TURBULNT	SENSIBLE
21	EXIT	1	1731.8	323.8	684.30	.000	1.9	.0	1574.567	128296.	TURBULNT	SENSIBLE
22	EXIT	2	1718.2	335.0	715.54	.000	1.9	.0	1574.567	141729.	TURBULNT	SENSIBLE
23	EXIT	3	1711.3	338.0	723.77	.000	1.9	.0	1574.567	145484.	TURBULNT	SENSIBLE
24	EXIT	4	1672.6	338.9	745.26	.012	4.3	98.3	1588.617	314426.	BUBBLY	TRANSITN
11	EXIT	5	1496.8	394.2	908.54	.015	5.0	94.5	1588.617	508580.	ANNULAR	TRANSITN
12	EXIT	6	1490.3	394.2	908.54	.015	5.0	94.3	1588.617	510123.	BUBBLY	SENSIBLE
13	EXIT	7	890.0	460.0	1119.53	.025	8.3	82.0	1588.617	1063298.	ANNULAR	NUCLEATE

Figure A.7 - Donges simulation results - Terminal conditions of each coil section at 130 ton/h feed rate

COIL SECT ID.	POINT INFO. AT	PROC. FLOW SEQ.	PRESSURE (KPAA)	TEMPERATURE (C)	ENTHALPY (J/G)	WEIGHT FRACT. VAPOR	VELOCITY (M/S)	CRITICAL VELOCITY (M/S)	MASS VELOCITY (KG/SEC/M2)	HOMOGEN. REYNOLDS NUMBER	FLOW REGIME	BOILING REGIME
21	INLET	1	2002.0	307.6	639.50	.000	2.3	.0	1937.929	135781.	TURBULNT	SENSIBLE
21	EXIT	1	1958.4	324.3	685.80	.000	2.4	.0	1937.929	158363.	TURBULNT	SENSIBLE
22	EXIT	2	1937.0	336.6	719.78	.000	2.4	.0	1937.929	176440.	TURBULNT	SENSIBLE
23	EXIT	3	1926.2	339.7	728.45	.000	2.4	.0	1937.929	181331.	TURBULNT	SENSIBLE
24	EXIT	4	1878.7	342.6	751.24	.010	4.5	104.1	1951.979	338433.	BUBBLY	TRANSITN
11	EXIT	5	1651.8	395.6	908.34	.012	5.2	97.5	1951.979	533528.	ANNULAR	TRANSITN
12	EXIT	6	1643.0	396.5	910.85	.012	5.2	97.2	1951.979	538124.	BUBBLY	NUCLEATE
13	EXIT	7	890.0	460.0	1115.58	.022	8.9	78.4	1951.979	1162012.	ANNULAR	NUCLEATE

Figure A.8 - Donges simulation results - Terminal conditions of each coil section at 160 ton/h feed rate

COIL SECT ID.	POINT INFO. AT	PROC. FLOW SEQ.	PRESSURE (KPAA)	TEMPERATURE (C)	ENTHALPY (J/G)	WEIGHT FRACT. VAPOR	VELOCITY (M/S)	CRITICAL VELOCITY (M/S)	MASS VELOCITY (KG/SEC/M2)	HOMOGEN. REYNOLDS NUMBER	FLOW REGIME	BOILING REGIME
21	INLET	1	2321.7	307.8	639.82	.000	2.9	.0	2422.412	169529.	TURBULNT	SENSIBLE
21	EXIT	1	2251.6	326.7	692.32	.000	3.0	.0	2422.412	201714.	TURBULNT	SENSIBLE
22	EXIT	2	2217.4	339.9	729.18	.000	3.0	.0	2422.412	226587.	TURBULNT	SENSIBLE
23	EXIT	3	2200.0	343.2	738.31	.000	3.0	.0	2422.412	233141.	TURBULNT	SENSIBLE
24	EXIT	4	2143.0	348.0	762.36	.008	4.8	115.0	2436.462	379403.	BUBBLY	TRANSITN
11	EXIT	5	1867.3	398.6	913.38	.009	5.4	104.9	2436.462	574006.	ANNULAR	TRANSITN
12	EXIT	6	1856.9	398.6	913.38	.009	5.4	104.5	2436.462	575806.	BUBBLY	SENSIBLE
13	EXIT	7	890.0	460.0	1112.11	.018	9.7	75.4	2436.462	1283403.	ANNULAR	NUCLEATE

Figure A.9 - Donges simulation results - Terminal conditions of each coil section at 200 ton/h feed rate