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Substitution of Fuels and Raw materials by Waste in High Temperature Processes

M. Beckmann, CUTEC-Institut GmbH & R. Scholz, Technical University Clausthal, Germany

Abstract

The energetic and material utilisation of waste has been the focus of much attention for a long time. A lot of references are found in literature with regard to the specific application e.g. the co-combustion of sewage sludge, the substitution of primary fuels with plastic wastes in the incineration process of the cement industry etc. Furthermore, comparisons of the individual applications with each other, as well as comparisons with state-of-the-art thermal waste treatment methods should be discussed. Naturally, when comparing these processes, the entirety of required additional material and energy expenditure must be included under equal boundary conditions. Here, among other things, special attention must be paid to the uniqueness of the respective case of application that results from the specific properties of the waste materials supplied. This holds especially true when examining the effects of replacing primary fuels with substitute fuels from waste (secondary fuels) e.g. –in the case of a combustion process– on heat transfer conditions, flow conditions and the related distribution of temperature, material transport and specific energy expenditure. Apart from material aspects, substitute fuels must also satisfy requirements pertaining to fuel- and heat-related aspects.

This article deals especially with demands concerning fuel- and heat-related aspects. For a better understanding, first of all general aspects of heat transfer in firings and industrial furnaces are explained. The so-called energy exchange ratio is important when evaluating a given case of fuel substitution. This ratio expresses the value of a substitute fuel in relation to that of the primary fuel and must be taken into consideration when the comparison is based on balancing equations. Moreover, by resorting to the example of the clinker process in the cement industry, this article addresses various consequences of fuel substitution on e.g. kiln temperatures, flue gas mass flows etc. from a process engineer's perspective. Using an example from the steel and iron industry, fuel substitution is considered for the case of a natural gas-fired, coke-free cupola.

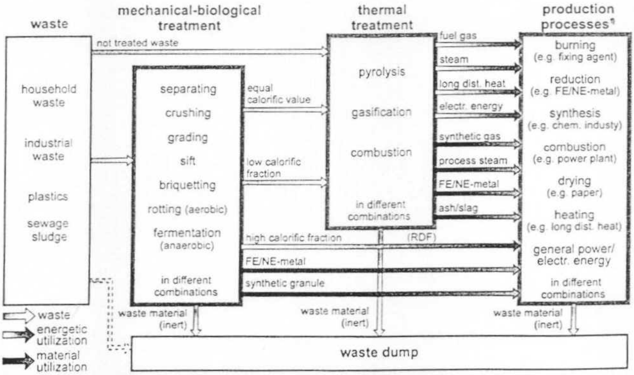
1 Introduction

Technical development in the field of waste management has lead to a situation today, in which several concepts are available for the treatment of waste. Increasingly, these also include the material and energetic utilisation of partial fractions in processes of material treatment or energy conversion.

Figure 1-1 illustrates these possibilities for domestic and industrial wastes, plastic wastes and sewage sludge. Classifying the options into the three process steps

- mechanical-biological treatment (pre-treatment),
- thermal waste treatment and
- production processes,

figure 1-1 confines itself to mentioning the respective basic operations. For each process step, further divisions are possible according to the different processes or, in the case of the production processes, according to the branch of industry. On this note, multiple references can be found in literature, for the mechanical-biological processes [e.g. 1 to 6], as



1) production integrated cycle economic is not regarded (top gas, dross, scrape from production, PMMA, coke)

Figure 1-1. Basic operations of waste treatment (Examples).

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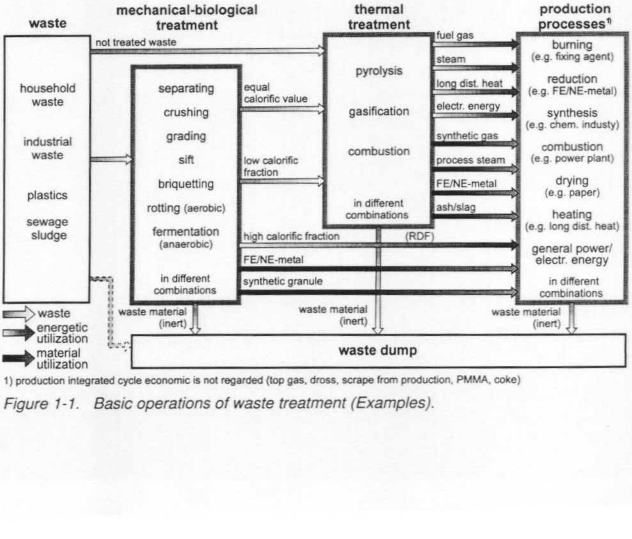

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well as for processes of thermal waste treatment [e.g. 7 to 15] and production processes [e.g. 16 to 23].

In figure 1-1 the arrows exemplifying material and energetic utilisation take into consideration

- not only the classical incineration of untreated waste (combustion-postcombustion processes with grate systems [e.g. 14], but also
- the generation of combustible gases by means of gasification, after having pre-treated the waste, e.g. for substituting primary fuels in material processing (calcinator stage in the clinker process [18]),
- the use of high calorific fuels (BRAM) and sewage sludge for co-combustion in power stations [23,24],
- the separation of plastics and metals for material utilisation [25,26]
- etc.

Moreover, figure 1-1 shows the need for investigating the entire course of the process from beginning to end after having dealt with a single process step. These investigations have been carried out for certain sorts of waste. For example, the energetic utilisation of plastic waste in the cement industry [27] or the treatment of domestic and industrial waste through mechanical-biological methods combined with thermal treatment processes deserve mention. Current research projects serve to diminish the deficits connected with the balancing of mechanical-biological treatment methods compared to thermal treatment methods and production processes. As a rule, the aforementioned comparisons are based on identical boundary conditions. Occasionally, this may require including additional reference processes or substitute processes. Furthermore, attention needs to be paid to process conditions changed by application of waste fractions in processes of material treatment and energy conversion as well as the thereby ensuing demand for appropriate optimisation of the process management.

Against this background, the following, first of all, investigates the general case of fuel substitution in processes of material treatment carried out in so-called industrial furnaces. Attention shall then be paid to the examples mentioned in the abstract, i.e. from the cement industry and steel and iron industry. Statements concerning fuel substitution can similarly be transferred to other processes of the raw material industry (lime, glass, ceramics etc.) as well as to processes of energy conversion in general, i.e. high temperature processes.

2 Balances of Firings and Industrial Furnaces including Heat Recovery

The term "industrial furnace" comprises all fossil fuel-fired or electrically heated apparatuses, in which solid, liquid or gaseous materials are subjected to thermal treatment on an industrial scale. In this context, the focus is on fossil fuel-fired industrial furnaces.

Industrial furnaces and firings for energy conversion have in common the process of combustion. However, compared to industrial furnaces where the combustion process is controlled in accordance with the load or input material to be treated, firings are optimised, e.g. in the case of steam generated in a boiler, principally with respect to the exploitation of energy. Figure 2-1 schematically illustrates this difference [28]. With regard to heat generation and exchange, the similarity between a boiler and an industrial furnace principally allows aspects of balancing, heat recovery and fuel substitution to be transferred from an example, treated in the following, of a material conversion process to energy conversion processes. In the case of directly fired industrial furnaces, fuel substitution is always linked to the question of product compatibility. Fuel with an ash content may then also require a respective material substitution.

Industrial furnaces are employed in almost all significant branches of the raw material industry. Examples include the burning of clinker in rotary kilns in the cement industry

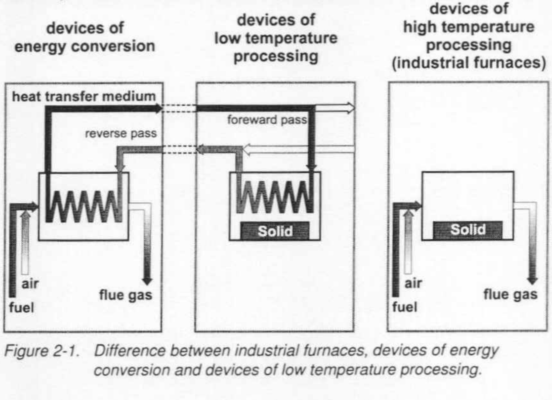

Figure 2-1. Difference between industrial furnaces, devices of energy conversion and devices of low temperature processing.

(figure 2-2), burning of lime stone in shaft furnaces, burning of ceramic products for the building industry in tunnel furnaces, generation of pig iron in blast furnaces, smelting of scrap in a cupola (figure 2-3) etc.

Commonly, energy balances form the basis for evaluating industrial furnaces, both for plant components as well as for the entire plant (preheating, combustion and cooling processes etc.). For a better understanding, figure 2-4 contains an example of air and fuel preheating and a continuous furnace process with a chemical or physical conversion process in the load (burning of lime, smelting of metals etc.). For the example in figure 2-4, the total balance reads:

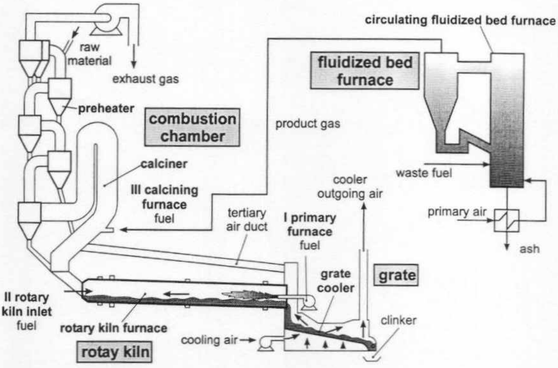

Figure 2-2. Schematic presentation of the cement process in cobination with a fluidized bed furnace.

$$\dot{m}_F \cdot h_n = \dot{m}_S \cdot \left[\Delta h_{RC} + c_{S21} \cdot (\vartheta_{S2} - \vartheta_{S1})\right] + \dot{m}_G \cdot c_{G3a} \cdot (\vartheta_{G3} - \vartheta_a) + \sum \dot{Q}_l \qquad (2.1).$$

In the general case depicted here, heat from the hot flue gas is exchanged in a recuperator to preheat the combustion air and, in a second recuperator connected in series, to preheat the fuel. Especially in the context of fuel substitution, the question arises, to what extent air and/or fuel preheating makes sense. The entering and exiting mass flows of air, fuel and gas ($\dot{m}_A, \dot{m}_F, \dot{m}_G$) needed for the balancing equations are connected with each other via the specific stoichiometric air requirement obtained through the combustion equation and by neglecting the solid inert components in the fuel (ash, e.q. (2.3)):

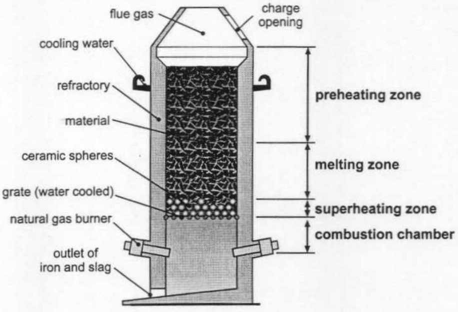

Figure 2-3. Schematic presentation of a natural gas fired, cokeless cupola furnace.

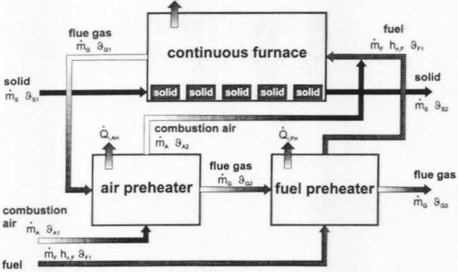

Figure 2-4. Schematic presentation of a continuous furnace with preheating of air and fuel.

$$\dot{m}_A = \lambda \cdot l_{min} \cdot \dot{m}_F \qquad (2.2)$$

$$\dot{m}_G = (1 + \lambda \cdot l_{min}) \cdot \dot{m}_F \qquad (2.3).$$

Thus, the efficiency factor of the air recuperator amounts to:

$$\eta_{AH} = \frac{\dot{m}_F \cdot \lambda \cdot l_{min} \cdot c_{A21} \cdot (\vartheta_{A2} - \vartheta_{A1})}{\dot{m}_F \cdot (1 + \lambda \cdot l_{min}) \cdot c_{G1a} \cdot (\vartheta_{G1} - \vartheta_a)} \qquad (2.4).$$

Accordingly, that of the fuel recuperator reads:

$$\eta_{FH} = \frac{\dot{m}_F \cdot c_{F21} \cdot (\vartheta_{F2} - \vartheta_{F1})}{\dot{m}_F \cdot (1 + \lambda \cdot l_{min}) \cdot c_{G2a} \cdot (\vartheta_{G2} - \vartheta_a)} \qquad (2.5).$$

As seen in eqs. (2.4) and (2.5), the heat capacity flow ratios of both recuperators are less than one. Thus, in a counterflow arrangement, air can be only heated to a maximum temperature of $\vartheta_{A2} = \vartheta_{G1}$ and fuel to a maximum temperature of $\vartheta_{F2} = \vartheta_{G2}$. Therefore, the maximum efficiency factor of air and fuel preheating results, with the simplified assumption that the heat capacities in eqs. (2.4) and (2.5) are, in pairs, approximately equal and for the limiting case of a stoichiometric plant operation with $\lambda = 1{,}0$ for $\vartheta_{A1} = \vartheta_{F1} = \vartheta_a$, in [29]:

$$\eta_{AH,max} = \frac{l_{min}}{1 + l_{min}} \qquad (2.6)$$

$$\eta_{FH,max} = \frac{1}{1 + l_{min}} \qquad (2.7).$$

The greatest possible overall efficiency factor of the heat recovery plant amounts, with eqs. (2.6) and (2.7) for the special case of adiabatic recuperators, to:

$$\eta_{HR,max} = \frac{1 + l_{min} + l_{min}^2}{\left(1 + l_{min}\right)^2} \qquad (2.8).$$

Figure 2-5 shows the maximum efficiency factors (eqs. (2.6) to (2.8)) over the stoichiometric air requirement. Improvement of $\eta_{HR,max}$ by fuel preheating is only achieved in the case of fuels with a low stoichiometric air requirement, thus, as a rule, only in the case with fuels of low calorific value. Peculiarities in the context of heat recovery must obviously be taken into consideration when replacing a primary fuel of high calorific value with a substitute fuel of low calorific value. The following section takes a closer look at this kind of fuel exchange.

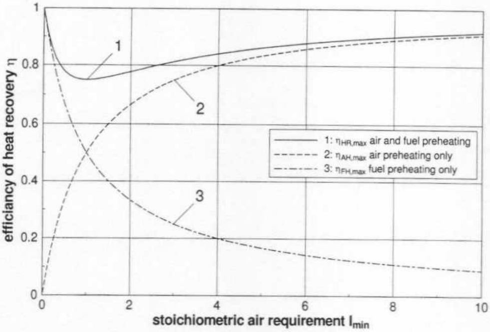

Figure 2-5. Maximum efficiencies for the heat recovery using preheating of air and/or fuel corresponding to the stoichoimetric air requirement of the fuel [29].

3 Potential of Fuel Substitution by Wastes

Usually, replacing fuels with waste will immediately lead to the question of the influence of the waste materials on the conditions of the respective process. Especially the effects of using substitute fuels will be examined on process temperatures, flue gas amounts, noxious substances or noxious matter loads and specific energy consumption in the case of industrial furnaces or, respectively, efficiency factors in the case of energy conversion plants. Not until then can possibilities be discussed of optimising the process with changed boundary conditions due to use of substitution. Thus, the assessment of a fuel depends not only on the kind of fuel itself, but it is also essentially influenced by the manner of operation of both plant and heat recovery.

Moreover, for the individual process, the ensuing mass, energy and material balances must be examined in the context of the entirety of the respective overall concept in the form of total balances [e.g. 15], i.e. among other things, the absolute figures of energy saved by means of substitution need to be compared with energy expenditures for required waste pre-treatment etc. The effects of substitution can essentially be derived from the combustion properties such as net calorific value h_n, stoichiometric or theoretical air requirement l_{min}, specific flue gas amount $v_{G,min}$, calorific combustion temperature ϑ_{cal} etc.

At first sight, the balancing equation (2.1), along with figure 2-4, shows that the flue gas loss of a given fuel is higher, the higher the amount of flue gas $\dot{m}_G$ in relation to the net calorific value h_n or the higher the flue gas temperature ϑ_{G3}. The increase of flue gas loss means less fuel energy can be utilised for the process, the specific energy expenditure rises.

Less apparent is the influence of the calorific combustion temperature on process conditions connected with the aforementioned quantities. For discussing this question, in a first step we examine one section of a continuous industrial furnace and – for the time being to explain the principle – for reasons of simplification liken it to a continuously stirred reactor element (CSR) (figure 3-1A, left). The supplied energy results from the conversion of fuel with air taking into account air and fuel preheating. For the time being, the basic considerations disregard the influence of a thermal dissociation balance and the theoretical combustion temperature ϑ_{th} that is determined accordingly. Thus, the caloric combustion temperature ϑ_{cal} can be used for obtaining the gas enthalpy flow. In an ideally mixed kiln chamber (CSR) this enthalpy flow effects a heat flow $\dot{Q}_{CSR}$ to the load with the con-

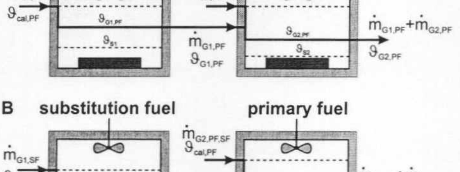

Figure 3-1. Heat transfer in industrial furnaces; model idea continuously stirred reactor elements.

stant surface temperature ϑ_{S1} and the temperature of the gas in the stirred reactor element settles at a constant temperature that equals the exit temperature ϑ_{G1}. From that, assuming adiabatic conditions, follows generally ($\vartheta_{S1} \equiv \vartheta_S$, $\vartheta_{G1} \equiv \vartheta_G$):

$$\Delta \dot{H}_G = \dot{Q}_{CSR} \qquad (3.1)$$

$$\text{with } \Delta \dot{H}_G = \dot{m}_G \cdot c_{G,calG} \cdot (\vartheta_{cal} - \vartheta_G) \qquad (3.2)$$

$$\text{and } \dot{Q}_{CSR} = \alpha_{\alpha\varepsilon} \cdot A_S \cdot (\vartheta_G - \vartheta_S) \qquad (3.3),$$

where in eq. (3.3) the heat transfer coefficient $\alpha_{\alpha\varepsilon}$ includes convection and radiation and A_S denotes the surface of the load.

When replacing the primary fuel (PF) with the substitute fuel (SF), the demand exists that the load temperature ϑ_S and the kiln output, i.e. the exchanged heat flow $\dot{Q}$, remain unchanged

$$\dot{Q}_{PF} = \dot{Q}_{SF} \qquad (3.4),$$

immediately yielding:

$$\Delta \dot{H}_{G,PF} = \Delta \dot{H}_{G,SF} \qquad (3.5).$$

Eq. (2.3) links the gas mass flows $\dot{m}_{G,PF}$ and $\dot{m}_{G,SF}$ from the fuel conversion with the respective fuel mass flows $\dot{m}_{PF}$ and $\dot{m}_{SF}$ while disregarding the solid inert component of the fuel (ash).

In introducing the energy exchange ratio E:

$$E = \frac{\dot{m}_{SF} \cdot h_{n,SF}}{\dot{m}_{PF} \cdot h_{n,PF}} \qquad (3.6),$$

which, from an energetic point of view, expresses the value of a substitute fuel in relation to the primary fuel, eq. (3.6) can be re-arranged to yield the ratio of fuel mass flows:

$$\frac{\dot{m}_{SF}}{\dot{m}_{PF}} = E \cdot \frac{h_{n,PF}}{h_{n,SF}} \qquad (3.7)$$

while eqs. (2.3) and (3.6) yield the ratios of flue gas flows:

$$\frac{\dot{m}_{G,SF}}{\dot{m}_{G,PF}} = E \cdot \frac{h_{n,PF} \cdot \left(1 + \lambda_{SF} \cdot l_{min,SF}\right)}{h_{n,SF} \cdot \left(1 + \lambda_{PF} \cdot l_{min,PF}\right)} \qquad (3.8).$$

The ratios of fuel and flue gas mass flows are important for further consideration of such boundary conditions that are linked to the plant itself such as fuel supply, flow conditions in the kiln chamber, material transport etc. Often these determine the maximum rate of substitution (replacement of primary with substitute fuel), which is dealt with at a later point.

Further detailed, with eqs. (2.3), (3.2) and (3.6), the energy exchange ratio for a continuously stirred reactor reads:

$$E_{CSR} = \frac{\left(1 + \lambda_{PF} \cdot l_{min,PF}\right) \cdot \dfrac{c_{G,PF,calG1} \cdot \left(\vartheta_{cal,PF} - \vartheta_{G1,PF}\right)}{h_{n,PF}}}{\left(1 + \lambda_{SF} \cdot l_{min,SF}\right) \cdot \dfrac{c_{G,SF,calG1} \cdot \left(\vartheta_{cal,SF} - \vartheta_{G1,SF}\right)}{h_{n,SF}}} \qquad (3.9).$$

During the following examinations we assume that the conditions of heat exchange $(\alpha_{\alpha\varepsilon} \cdot A_S)$ are not influenced by the substitution (static examination). At a later point, the influence of changed conditions of heat transfer is separately dealt with in an advanced step. From the static examination assumed for the time being follows:

$$\vartheta_{G1,SF} = \vartheta_{G1,PF} \qquad (3.10).$$

With the accepted simplifications, the energy exchange ratio $E_{CSR,\vartheta}$ can be obtained for a given replacement of a primary fuel with a substitute fuel.

As a rule, substitute fuels with lower calorific values than the primary fuels also achieve lower calorific combustion temperatures. Therefore, when employing substitute fuels, in order to balance the smaller difference of potential between the calorific temperature and the balancing or gas temperature (ϑ_{cal} - ϑ_G), compared to that of the primary fuel, the gas mass flow $\dot{m}_G$ must increase accordingly. Thus, compared to the primary fuel, the use of substitute fuel at first entails a higher total energy expenditure. The energy exchange ratio expresses the ratio between both total energy expenditures. In the context of substitute fuel employment, the aim of optimisation is, by appropriate plant arrangements and heat recovery etc., to attain a smallest possible energy exchange ratio. This is explained in further detail at a later point.

Figure 3-2 shows $E_{CSR,\vartheta}$ according to eq. (3.9) over a required gas temperature ϑ_G for the case of replacing a primary fuel that has a calorific value of $h_{n,PF}$ = 30 MJ/kg with a substitute fuel with $h_{n,SF}$ = 11, 15, 35 and 45 MJ/kg. The energy exchange ratio increases, the higher the required balancing temperature and the lower the calorific value of the substitute fuel as compared to the primary fuel (curves 1 to 4 in figure 3-2).

Conversely, in the case of the substitute fuel having a higher calorific value than the primary fuel, the energy exchange ratio assumes values less than one (curves 5 to 8). This, for example, is the case when replacing coal with high calorific plastic wastes in the primary firing of the clinker burning process (cf. figure 2-2). When comparing the energy exchange ratios in figure 2-2 for a given substitute fuel with and without air preheating, the significance of heat recovery (e.g. recovery of heat from the load in the grate cooling-clinker burning process) becomes evident. In the case of a substitute fuel with $h_{n,SF}$ = 11 MJ/kg (curves 1 and 2), gas temperatures $\vartheta_G > 1700$ °C can only be achieved with appropriate preheating of air.

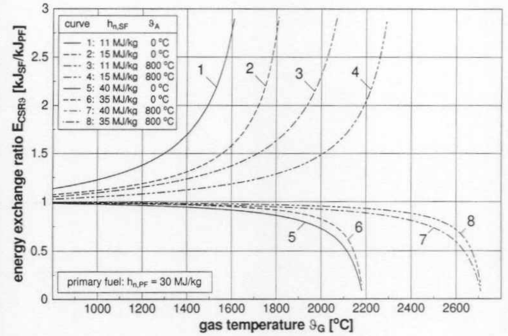

Figure 3-2. Energy exchange ratio by static observation in dependence of the gas temperature for different substitution cases.

Until now, the gas temperature ϑ_G with substitution was assumed constant. However, due to changed mass flows and compositions of the gas, fuel replacement also leads to changed conditions of heat transfer (convection and radiation), thus, varying gas temperatures ($\vartheta_{G,PF} \neq \vartheta_{G,SF}$) are attained. The following considerations refer to the case of an exclusive transfer of heat by means of convection. The general case reads:

$$\dot{Q}_{CSR,\alpha} = \alpha \cdot A_S \cdot (\vartheta_G - \vartheta_S) \qquad (3.11).$$

With Stanton's number:

$$St = \frac{\alpha \cdot A_S}{\dot{m}_G \cdot c_G} = \frac{(\vartheta_{cal} - \vartheta_G)}{(\vartheta_G - \vartheta_S)} \qquad (3.12),$$

the transferred heat flow becomes

$$\dot{Q}_{CSR,\alpha} = \dot{m}_G \cdot c_G \cdot \frac{1}{1 + 1/St} \cdot (\vartheta_{cal} - \vartheta_S) \qquad (3.13).$$

Hence, the energy exchange ratio $E_{CSR,\alpha}$ is expressed by the respective fuel-related quantities h_n, l_{min}, ϑ_{cal}, the process-related, fixed temperature of the load ϑ_S, and finally by Stanton's number, the latter characterising the heat transfer conditions:

$$E_{CSR,\alpha} = \frac{(1 + \lambda_{PF} \cdot l_{min,PF}) \cdot c_{G,PF} \cdot \dfrac{1}{\left(1 + \dfrac{1}{St_{PF}}\right)} \cdot \dfrac{(\vartheta_{cal,PF} - \vartheta_S)}{h_{n,PF}}}{(1 + \lambda_{SF} \cdot l_{min,SF}) \cdot c_{G,SF} \cdot \dfrac{1}{\left(1 + \dfrac{1}{St_{SF}}\right)} \cdot \dfrac{(\vartheta_{cal,SF} - \vartheta_S)}{h_{n,SF}}} \qquad (3.14).$$

In order to evaluate eq. (3.14), St_{PF} and St_{SF} require further examination.

Heat transfer by convection is commonly described by functions of the following kind:

$$Nu = k \cdot Re^a \cdot Pr^b \cdot \left(\frac{L_K}{L}\right)^c \qquad (3.15)$$

With a = 0,8 and assuming constant material-related quantities, the ratio of Stanton numbers reads:

$$\frac{St_{PF}}{St_{SF}} = \left(\frac{\dot{m}_{G,SF}}{\dot{m}_{G,PF}}\right)^{0,2} = \left(E_{CSR,\alpha} \cdot \frac{h_{n,PF} \cdot (1 + \lambda_{SF} \cdot l_{min,SF})}{h_{n,SF} \cdot (1 + \lambda_{PF} \cdot l_{min,PF})}\right)^{0,2} \qquad (3.16).$$

As a rule, a vast amount of data concerning operation with primary fuel is available, allowing the related Stanton numbers St_{PF} to be calculated. Experience shows that, in the case of the high temperature applications especially dealt with here, the Stanton numbers range from St_{PF} = 2 to St_{PF} = 4. This stipulation enables the calculation of an energy exchange ratio that takes into account the means of heat exchange –here, convection– $E_{CSR,\alpha}$, with eqs. (3.14) and (3.16).

Figure 3-3 presents $E_{CSR,\alpha}$ over the ratio of calorific values $h_{n,SF}/h_{n,PF}$ for $St_{PF} = 2$ and a required temperature of the solid of $\vartheta_S = 1500$ °C as well as with air preheating ($\vartheta_A = 800$°C) as compared to the previously discussed $E_{CSR,\vartheta}$ for the static examination. The boundary conditions for this example roughly resemble those of the combustion zone of a clinker burning process (primary firing) or those of a smelting processes in the steel and iron industry. Here, a comparison of the two curves for $E_{CSR,\vartheta}$ and $E_{CSR,\alpha}$ shows that consideration of changed heat transfer conditions by substitution leads to $E_{CSR,\alpha} < E_{CSR,\vartheta}$. However, only heat exchange by convection was assumed. Similarly, heat exchange by radiation can be included. In this case, the Konakov number takes the place of Stanton's number, stating the ratio between heat capacity flow of the gas and the heat flow transferred by radiation. In the framework given here we will refrain from going into further detail concerning radiation, however, it should be mentioned that a deterioration of the radiation properties of the gas from substitute fuel combustion is possible too. This could lead to less favourable conditions for heat exchange (i.e. $E_{CSR,\alpha} > E_{CSR,\vartheta}$) when compared to the use of primary fuel, thus resulting in possible gas temperatures $\vartheta_{G,SF} > \vartheta_{G,PF}$.

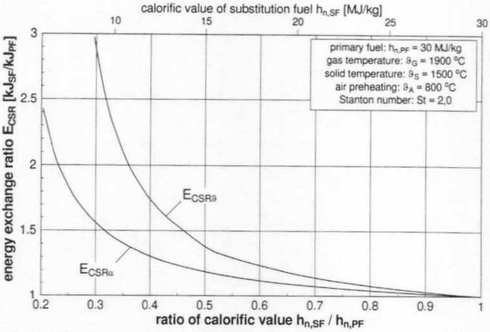

Figure 3-3. Energy exchange ratio by static observation in dependence of the ratio of calorific values for different heat transfer setups.

In the first step discussed above, only one section of an industrial furnace was examined under simplified conditions (just one continuously stirred reactor element) with regard to the energy exchange ratio of fuel substitution. However, as mentioned above, the evaluation of a fuel depends not alone on the kind of fuel but also on process management and heat recovery. Therefore, the amount of additional energy expenditure in the case of fuel substitution, as opposed to primary fuel use, required primarily for attaining high process temperatures, e.g. in a burning or calcining process, is not necessarily "lost".

For example, with a two-fold staging of the energy input along the treatment path (primary and secondary firing with the clinker burning process), a balance schema as in figure 3-1 can be derived, again making use of the aforementioned simplified model of continuously stirred reactor elements for the first and second stage and assuming static examination ($\vartheta_{G,PF} = \vartheta_{G,SF}$). The single figures 3-1A present the first and second stage of the original process, i.e. with primary fuel supply in the respective stages ($\dot{m}_{PF1}$) and ($\dot{m}_{PF2,PF}$). This process is now compared to a case in which, in the first stage, the primary fuel supply is replaced with substitute fuel $\dot{m}_{SF1}$ according to the exchange ratio $E_{CSR,\vartheta}$ mentioned above (figure 3-1B). For the second stage the same requirement applies that does to the first stage, namely, that despite fuel substitution, the temperature of the load ϑ_{S2} and kiln output remain the same (eqs. (3.4) and (3.5)). When employing substitute fuel, the higher amount of energy (E_{CSR1}) supplied to the first stirred reactor element, in comparison to the use of primary fuel, now can be utilised with the amount that lies above the required temperature level of the second stage, and, accordingly, less primary energy needs to be supplied to the second stage (second continuously stirred reactor). The energy balance for the second stage yields with $\vartheta_{G1,PF} = \vartheta_{G1,SF} = \vartheta_{G1}$ and $\vartheta_{G2,PF} = \vartheta_{G2,SF,PF} = \vartheta_{G2}$ (static examination):

$$\begin{aligned}
&\left(1+\lambda_{PF}\cdot l_{min,PF}\right)\cdot \dot{m}_{PF1}\cdot c_{G1,PF,G1G2}\cdot\left(\vartheta_{G1}-\vartheta_{G2}\right)+\\
&\left(1+\lambda_{PF}\cdot l_{min,PF}\right)\cdot \dot{m}_{PF2,PF}\cdot c_{G2,PF,calG2}\cdot\left(\vartheta_{cal,PF}-\vartheta_{G2}\right)=\\
&\left(1+\lambda_{SF}\cdot l_{min,SF}\right)\cdot \dot{m}_{SF1}\cdot c_{G1,SF,G1G2}\cdot\left(\vartheta_{G1}-\vartheta_{G2}\right)+\\
&\left(1+\lambda_{PF}\cdot l_{min,PF}\right)\cdot \dot{m}_{PF2,SF}\cdot c_{G2,PF,calG2}\cdot\left(\vartheta_{cal,PF}-\vartheta_{G2}\right)
\end{aligned} \qquad (3.17).$$

From eqs. (3.7) and (3.9), we obtain the difference of primary fuel mass flows in the second stage with and without substitution in relation to the original primary fuel mass flow in stage 1:

$$\Delta x = \frac{\dot{m}_{PF2,PF} - \dot{m}_{PF2,SF}}{\dot{m}_{PF1}} =$$

$$\frac{c_{G1,PF,calG1} \cdot \left(\vartheta_{cal,PF} - \vartheta_{G1}\right) \cdot c_{G1,PF,G1G2} \cdot \left(\vartheta_{G1} - \vartheta_{G2}\right)}{c_{G1,SF,calG1} \cdot \left(\vartheta_{cal,SF} - \vartheta_{G1}\right) \cdot c_{G2,PF,calG2} \cdot \left(\vartheta_{cal,PF} - \vartheta_{G2}\right)} - \frac{c_{G1,PF,G1G2} \cdot \left(\vartheta_{G1} - \vartheta_{G2}\right)}{c_{G2,PF,calG2} \cdot \left(\vartheta_{cal,PF} - \vartheta_{G2}\right)} \qquad (3.18).$$

With the ratio of fuel distribution between first and second stage in the case of exclusive input of primary fuel

$$y = \frac{\dot{m}_{PF2,PF}}{\dot{m}_{PF1} + \dot{m}_{PF2,PF}} \qquad (3.19)$$

the energy exchange ratio for the second stage (static examination) can be obtained from eq. (3.18):

$$E_{CSR2,\vartheta} = \frac{\dot{m}_{PF2,SF}}{\dot{m}_{PF2,PF}} = \Delta x \cdot \left(1 - \frac{1}{y}\right) + 1 \qquad (3.20).$$

$E_{CSR,\vartheta}$ depends on the respective temperature differences and the original fuel distribution y. Generating the energy exchange ratio for both stirred reactor elements:

$$E_{CSR12,\vartheta} = \frac{\dot{m}_{SF1} \cdot h_{n,SF} + \dot{m}_{PF2,SF} \cdot h_{n,PF}}{\left(\dot{m}_{PF1} + \dot{m}_{PF2,PF}\right) \cdot h_{n,PF}} \qquad (3.21)$$

leads with eqs. (3.6), (3.19) and (3.20), as expected, to

$$E_{CSR12,\vartheta} = (1 - y) \cdot E_{CSR1,\vartheta} + y \cdot E_{CSR2,\vartheta} \qquad (3.22).$$

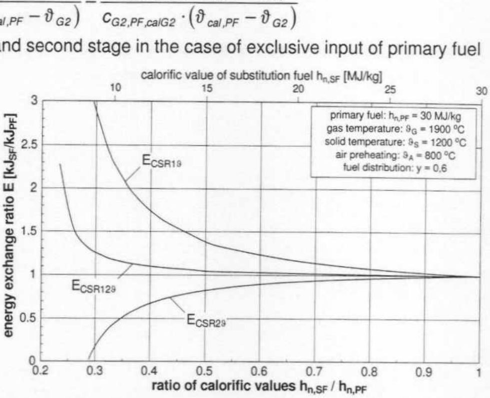

Figure 3-4. Energy exchange ratio by static observation in dependence of the ratio of calorific values for two stage fuel feed.

As in the case of the previous example, with a primary fuel that has $h_{n,PF}$ = 30 MJ/kg, a balancing temperature ϑ_{G1} = 1900 °C and air preheated to ϑ_A = 800 °C, figure 3-4 presents the energy exchange ratio $E_{CSR1,\vartheta}$ over $h_{n,SF}$.

Furthermore, for a fixed balancing temperature ϑ_{G2} = 1200 °C, the curve for $E_{CSR2,\vartheta}$ is included, the course of which, due to the savings achieved in the second continuously stirred reactor, assumes values less than 1 in this presented substitution case. Notwithstanding the fact that $E_{CSR12,\vartheta}$ is greater than 1, due to the possibility of a staged fuel supply, it is still significantly smaller than $E_{CSR1,\vartheta}$.

In the same way as with a single continuously stirred reactor element, the heat transfer conditions can be included for two, namely, by connecting them in series. Moreover, with a cascade of stirred reactors, the residence time of the real process can be approximated. However, for a number of industrial furnaces this can also be adequately described by plug flow reactor (PFR) characteristics. In this case, relations can be called on that are known from continuous heat exchange systems. Under adiabatic conditions, the energy exchange ratio of a plug flow element (single stage) is described by:

$$E_{PFR} = \frac{c_{G,PF,calG} \cdot \left(\vartheta_{cal,PF} - \vartheta_{G,PF}\right)}{c_{G,SF,calG} \cdot \left(\vartheta_{cal,SF} - \vartheta_{G,SF}\right)} \cdot \frac{c_{G,SF,kala} \cdot \left(\vartheta_{cal,SF} - \vartheta_a\right)}{c_{G,PF,kala} \cdot \left(\vartheta_{cal,PF} - \vartheta_a\right)} \approx \frac{1 - \dfrac{\vartheta_{G,PF} - \vartheta_a}{\vartheta_{cal,PF} - \vartheta_a}}{1 - \dfrac{\vartheta_{G,SF} - \vartheta_a}{\vartheta_{cal,SF} - \vartheta_a}} \qquad (3.23).$$

The simplification of eq. (3.23), in which the specific heat capacities have been disregarded, shows that, when assuming constant exiting temperatures $\vartheta_{G,PF/SF}$ for the time being, a decrease of the substitute fuel's calorific combustion temperature $\vartheta_{cal,SF}$ compared to $\vartheta_{cal,PF}$, results in an energy exchange ratio E_{PFR} with values greater than one. When, due to the respective heat transfer conditions, with employment of substitutes fuels the exiting temperatures settle at $\vartheta_{G,SF} > \vartheta_{G,PF}$, E_{PFR} is caused to increase further, while conversely, when $\vartheta_{G,SF} < \vartheta_{G,PF}$, the result is a decrease accordingly.

4 Effects of Substitution on Process Management (Examples)

With the above acquired and examined findings, greatly simplified with regard to some general aspects of fuel substitution, the following examples are discussed concerning the impact of fuel substitution on the respective process management:

- clinker burning process in the cement industry and
- smelting process in a cupola in the steel and iron industry.

The underlying presumption in the following considerations is that the substitute fuels, based on their material properties, are principally suitable for substitution both with regard to product quality as well as to emission of pollutants [e.g. 30]. In Germany, cement clinker is produced mainly according to the so-called dry method. As seen in figure 2-2, the plant consists of the main components preheater, calcinator (combustion chamber), rotary kiln and clinker cooler (grate). Fuel is supplied through three inlets: the primary firing (kiln exit, I in figure 2-1), the secondary firing (calcinator, II) and the supplemental firing at the kiln entrance (III). In principle, substitute fuel can be supplied at all three inlets, however, the specific boundary conditions need to be adhered to.

In the combustion zone, the primary firing for the burning of the clinker needs to achieve approximate maximum load temperatures around $\vartheta_S \approx 1500$ °C or gas temperatures $\vartheta_G \approx 1900$ °C (cf. figure 2-1, temperature curve). For example, a question could be, to what extent coal supplied as primary fuel with $h_{n,PF} = 30$ MJ/kg can be replaced with a substitute fuel gained from adequately pre-treated waste with $h_{n,SF} = 15$ MJ/kg. In this case, the energy exchange ratio, for the time being, only needs to be examined for the first stage. With the simplified relations derived in chapter 3, figure 3-2 states $E'_{CSR,\vartheta} \approx 1{,}3$ for this case. This entails a 2.6-fold amount of fuel and about 1.5-fold amount of flue gas when using the substitute fuel as opposed to the primary fuel (eqs.(3.7) and (3.8)). These circumstances prohibit a complete substitution, amongst other things with regard to aspects concerning the plant such as supply units, gas velocities in the rotary kiln, sweeping away of particles of the load by the gas flow etc. Therefore, with regard to justifiable changes of the boundary conditions, only amounts of 30 to 50 mass-% could be replaced in this case. With a primary fuel of $h_n = 30$ MJ/kg, a 50 % exchange with substitute fuel would result in a new mean calorific value of $h_{n,SF} = 22{,}5$ MJ/kg. Thus, the energy exchange ratio amounts to a mere $E_1 \approx 1{,}1$.

In the second step, the effects need to be investigated of the substitution in the primary firing on the secondary firing that is still being operated on coal. As previously mentioned, this staged manner of supplying fuel enables a part of the additional energy expenditure, due to substitution and required for the incineration process, to be utilised in the succeeding sections. With cement rotary kiln plants, the share of fuel in the second firing commonly ranges at about $y = 0.6$. Due to the limitation of the substitution rate in the first stage, the consequences for the overall process are low. For the example observed here, a total enegy exchange ratio near one is obtained. Therefore, depending on the process management, a smaller overall energy exchange ratio can be obtained. Beyond the scope of the simplified relations presented here the determination of this ratio, however, requires suitable process models and practical research. As a rule, without additional measures for heat recovery, replacement with a substitute fuel of lower calorific value entails an increase of the specific energy expenditure. For example, [31] reports of practical experience with a change-over of an incineration process from fuel oil ($h_n \approx 40$ MJ/kg) to lignite dust ($h_n \approx 20$ MJ/kg). Amongst other things, this modification lead to a greater flame length, a shift of the sintering zone toward the kiln entrance and to a temperature rise at the kiln entrance as well as behind the preheater. Naturally, here, the differing burnout behaviour of lignite dust and fuel oil plays a part. Altogether, the outcome was an increase of specific energy expenditure by about 170 kJ/kg_{cl}. The additional expenditure is attributed in equal parts to the increase of the flue gas mass flow on the one hand and to the fluctuation of grain sizes – and its respective effect on the course of temperature – on the other hand.

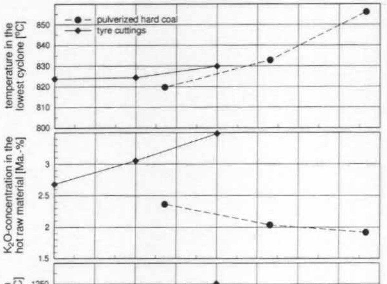

Figure 4-1. Influence of secondary fuel feeding on the process behaviour.

Concerning the secondary firing, due to lower temperatures compared to the primary firing, the energy exchange ratio in relation to the temperature is of less significance. According to figure 3-2, required gas temperatures of $\vartheta_G \approx 1200$ °C entail energy exchange ratios only slightly greater than one, even when the calorific value is lowered from 30 MJ/kg to 15 MJ/kg. Of significance is, when applying substitution to the secondary firing, the quality of the fuel with regard to the specific flue gas volume and the burnout. Larger-size particles such as tyre cuttings fall into the rotary kiln's entrance area where they cause a rise in temperature. Through incomplete combustion, the process of incineration could extend into the preheating zone. Concerning this, figure 4-1 displays results from practical research [32]. Amongst other things, the figure shows that, when applying tyre cuttings, the increase of secondary fuel energy in relation to the clinker mass causes the gas temperatures to rise at the kiln entrance. This temperature rise in turn effects an increased alkaline salt vaporisation in the rotary kiln. Conversely, when using a fuel with relatively good burnout properties, the conversion takes place mainly in the calcinator's gas rising duct and a reduction of alkaline salt recirculation can be expected. In the present example, an increase of the secondary fuel share through input of coal dust causes, due to a reduction in the primary fuel share, a fall in gas temperature at the kiln's entrance, thereby reducing the release of K_2O. A reduced amount of alkalines in the cycle finally leads to a smaller specific energy expenditure. Here, however, a complete burnout of the secondary fuel in the calcinator is presumed.

A further example for illustrating energy exchange ratios, the following examines the replacement of natural gas with gaseous substitute fuels of lower calorific value in the process of smelting cast iron, cast iron scrap etc. in a coke-free, natural gas-fired cupola [33,34]. As seen in figure 2-3, the object of interest consists of a counter-current-operated shaft furnace where the load proceeds in descending motion. In batches, i.e. discontinually, the load (pig iron, cast iron scrap, steel scrap) is supplied to the kiln through the kiln's feeding opening and is then preheated to melting point in the first zone. Adjacent to the preheating zone is the melting zone and –in order to prevent the melt solidifying around the water cooled grate bars– the superheating zone, the latter consisting of ceramic balls. In the combustion chamber natural gas is burnt with air and a supplemental supply of oxygen, altogether in excess of oxygen ($\lambda > 1$), thereby obtaining gas temperatures of 2100 °C to 2200 °C. Moreover, a burnout of the load takes place in the combustion chamber as well as in the shaft, in the magnitude of about 15 % of the overall heat input. Superheated to a temperature of about 1400 °C, the molten iron continually flows via a siphon with slag extraction through a channel for further processing [33]. Known since the late sixties, the coke-free, natural gas-fired cupola presents, from an economical and ecological point of view, an attractive alternative to the coke cupola, however, amongst other things due to lack of knowledge from a process engineering point of view, it lacked the resources to assert itself on an industrial scale [34, 25]. For describing the smelting behaviour of a coke-free, natural gas-fired cupola, an energetic process model has been developed that by now dependably reproduces the complex, inter-related processes of the smelting operation, as a comparison with experimental results confirms [33]. Figure 4-2 contains the calculated course of temperature for both material and flue gas. This process model also allows an examination of the subject of fuel substitution. The principle explanation of the energy exchange ratio in section 3 is based on greatly simplified relations. The aforementioned process model, however, contains the approaches regarding heat transfer in a much more complex form. When using this process model for examining the effects of fuel substitution, the generally explained finding is confirmed, stating that a decreasing calorific value of the applied substitute fuels entails an increase in the energy exchange ratio (figure 4-3). In contrast to the previous example of a clinker burning process, a cupola does not permit a second stage compensation of the heat capacity flow, increased due to fuel substitution. The temperatures and mass flows of gas leaving the preheating zone,

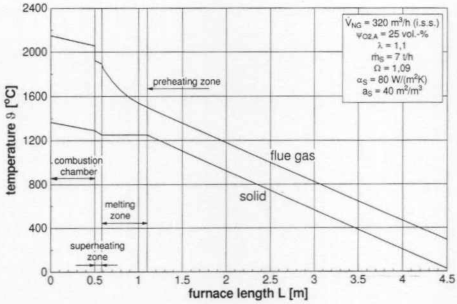

Figure 4-2. Calculated temperature curve of material and flue gas in a natural gas firded, cokeless cupola furnace.

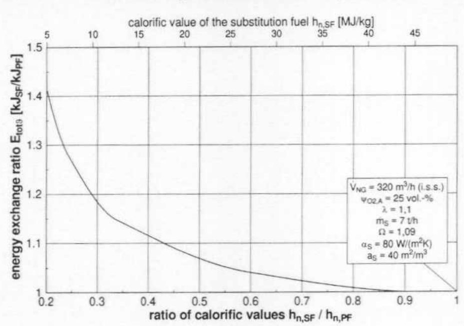

Figure 4-3. Energy exchange ratio by static observation in dependence of the fuel substitution for the cupola furnace.

designed as a counter-current heat exchanger, increase in the case of substitution which, accordingly, means a rise in losses and energy exchange ratio. Of course the thought occurs of, in the case of fuel substitution, utilising the flue gas flow behind the preheating zone for further heat recovery (e.g. for air preheating). However, apart from the aspects described in this context in section 2 (eq. (2.6)), attention needs to paid to the fact that the flue gases leaving the preheating zone contain a relatively high proportion of dust. Similar to the management of the cement process – where, in the case of substitution, the ratio of heat capacity flows in the counter-current preheating zone is compensated by a respective smaller input of secondary fuel – in the present example of the cupola it seems worth considering extracting a partial flow from the flue gas flow behind the smelting zone with ensuing heat recovery in order to improve the energy exchange ratio.

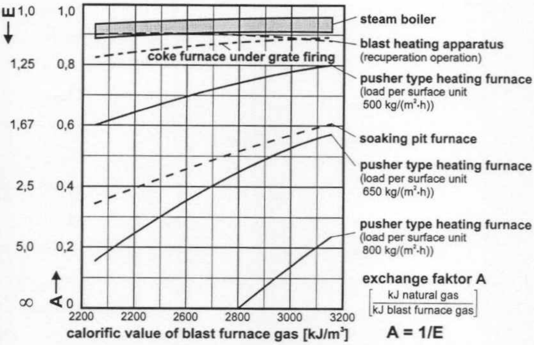

Figure 4-4. Substitution of natural gas by blast furnace gas by different consumers.

On a final note, in the context of the integrated economy of a steel plant, the energy exchange ratio is presented in figure 4-4 (as a factor $A = 1/E$), for the case of substituting natural gas with blast furnace gas, summarising a variety of processes [36]. When using blast furnace gas in the hot blast stoves (cowpers), boilers and furnaces heated from below of coke ovens, an exchange factor or value factor in relation to natural gas of 85 % to 90 % can be obtained. Applying blast furnace gas in reheating furnaces, i.e. processes with comparatively higher temperatures, the value factor attains a mere 30 % to 50 %. From this value factor certain strategies for plant conception and the use of so-called couple energy carriers can be derived. Therefore, in individual cases of fuel substitution, the need exists for examining which overall concept offers advantages with regard to energy savings.

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6 Nomenclature

Symbols

A	area; exchange factor
a	specific surface area
c	specific heat capacity (in connection with temperature differences always mean specific heat capacity)
E	energy exchange ratio
h	specific enthalpy
H	enthalpy
i.s.s.	in standard state
k	constant
l	specific air requirement
L	length
m	mass
Q	heat
v	specific volume
V	volume
x	stage 2 fuel mass in relation to stage 1
y	stage 2 fuel mass in relation to total
α	heat transfer coefficient
Δ	difference
η	efficiency factor
λ	stoichiometric ratio; air ratio
ϑ	temperature [°C]
ψ	concentration (volume related)
Nu	Nusselt number
Pr	Prandtl number
Re	Reynolds number
St	Stanton number

Indices (in superscript)

a,b,c	exponents
·	flow

Indices (in subscript)

a	ambient
A	air
AH	air preheating
cal	calorific
CF	continuous furnace
cl	clinker
CSR	continuously stirred reactor
F	fuel
FH	fuel preheating
G	flue gas
HR	heat recovery
l	loss
max	maximum
min	minimum
n	net (calorific value)
NG	natural gas
O_2	oxygen
PF	primary fuel
PFR	plug flow reactor
RC	reaction/material conversion
S	solid
SF	substitute fuel
th	theoretical
$1,2$	stages 1, 2
α	convection
Δ	difference
ε	radiation
ϑ	constant gas temperature (static examination)