Prospects and physical limits of processes and technologies in glass melting

Reinhard Conradt

To cite this article: Reinhard Conradt (2019): Prospects and physical limits of processes and technologies in glass melting, Journal of Asian Ceramic Societies, DOI: 10.1080/21870764.2019.1656360

To link to this article: https://doi.org/10.1080/21870764.2019.1656360

© 2019 The Author(s). Published by Informa UK Limited, trading as Taylor & Francis Group on behalf of The Korean Ceramic Society and The Ceramic Society of Japan.

Published online: 04 Sep 2019.
Prospects and physical limits of processes and technologies in glass melting

Reinhard Conradt

emeritus Chair of Glass an Ceramic Composites, RWTH Aachen University, Aachen, Germany; uniglassAC GmbH, Aachen, Germany

ABSTRACT

The paper is devoted to the physics of continuously operating fuel-fired glass furnaces with supplementary electrical boosting. Furnaces are treated in their basic function as heat exchangers and chemical reactors. First, as an expression of the 1st law of thermodynamics, a general heat balance is elaborated in detail. The function as heat exchanger is characterized by three dimensionless key indicators: the temperature efficiency (comprising adiabatic flame, glass exit, and environmental temperature), the heat capacity flow match of hot and cold streams, and the number of heat transfer units. A 2nd law treatment reveals the bottle neck of furnace optimization between two conflicting objectives, i.e., high production rates and high energy efficiency. Based on this treatment, an evaluation procedure for furnace performance is presented. It rests on a retrospective analysis of furnace operation data and allows one to quantitatively compare furnaces of different sizes and production capacities as well as the effect of different batches used in the same furnace. A number of industrial case studies demonstrate the usefulness and reliability of the approach. Finally, an expression for the ultimate physical limit of energy utilization efficiency of a fuel fired furnace in general is derived.

1. Introduction

Among the heavy industries performing high-temperature processes, the glass industry plays an important role worldwide. Measured the amount of glass produced, 45%, 28%, 12% and 15% go to the container, flat, fiber (insulation and reinforcement) and specialty glass branch, respectively. Since the oil crisis in the early 1970s, the issue of energy utilization has become a major concern. In recent years, the climate change debate with its focus on the release of CO2 into the atmosphere – besides the emission of hazardous species such as NOx – has strongly intensified the situation. The amount of released CO2 is chiefly due to the still dominant use of fossil fuel in the melting process. In fact, the glass industry has been striving for a reduction in energy demand since the late 19th century. The ground-breaking innovation at that time was the invention of the regenerative furnace by Sir Karl Wilhelm Siemens, younger brother of Werner Siemens. Much like some other inventions emerging at that time, i.e. the Otto and the Diesel engine, this concept was so outstandingly efficient that it has dominated the design of glass furnaces up to the present. Historically, energy demand for the production of 1 t of glass has decreased as follows:

- 20200 MJ/t in 1930,
- 9400 MJ/t in 1970,
- 5800 MJ/t in 1990.

Throughout the introduction, energy demand is given in MJ/t. In the main body, the kWh/t unit is chosen to allow for an easier comparison to electrical energy. Units are converted as follows:

- 1 kWh = 3600 MJ = 860.4 kcal = 3412.1 BTU;
- 1 t = 1000 kg = 1.1023 ton° where ton° denotes short tons.

Thus, 1 MBTU/ton° is equivalent to 1163 MJ/t.

The earlier decrease (1930–1970) in the above compilation was due chiefly to progress in the field of refractory materials allowing the use of increased processing temperatures. While a glass furnace in 1930 had a lifetime as short as just 1 to 1.5 years, lifetimes well beyond 10 years are now standard. The significant further decrease (1970–1990) was due mainly to systematic use of recycled cullet. Depending on the glass color, 30% to 70% and often up to 90% of cullet is used in the container glass category while 25% is a typical level for flat glass. Thus, both the development of refractory materials and the use of cullet must be considered as the two major milestones in energy saving. Lately [1,2], a benchmark survey on >130 container glass furnaces was presented. Energies were normalized to 50% cullet and to the primary energy equivalent of oxygen production for oxy-fuel furnaces. The ranking yielded the following:

- 3850 MJ/t for the best furnace,
- 4300 MJ/t for the top 10% performers,
- 5150 MJ/t on average.

CONTACT Reinhard Conradt reinhard.conradt@gmail.com uniglassAC GmbH, Nizzaallee 75, Aachen, Germany

© 2019 The Author(s). Published by Informa UK Limited, trading as Taylor & Francis Group on behalf of The Korean Ceramic Society and The Ceramic Society of Japan. This is an Open Access article distributed under the terms of the Creative Commons Attribution License (http://creativecommons.org/licenses/by/4.0/), which permits unrestricted use, distribution, and reproduction in any medium, provided the original work is properly cited.
In a recent update [3], the top 10% performers reached 3900 MJ/t. Obviously, the number of furnaces with good performance increased, while the lower limit did not shift. In an own study performed on a smaller lot (<30 cases), the best furnace reached 3750 MJ/t. This is consistent with other investigations [4] yielding best cases, which achieved 3700–3800 MJ/t. Additional batch pre-heating resulted in 3670 MJ/t [5]. All-electric furnaces [6] were reported to reach 3100–3200 MJ/t. All values are normalized as before. For float glass furnaces using 25% cullet, the energy demand essentially depends on the furnace size, and hence, on the production rate; the following typical values are maintained [1]:

- 7250 MJ/t at 600 t/d,
- 6500 MJ/t at 800 t/d,
- 5000 MJ/t at 1000 t/d.

The historical development of the data compiled above suggests that the physical limits of the performance of conventional furnaces is undoubtedly approached.

In the efforts toward rendering glass furnaces, more energy efficient, countless evolutionary and revolutionary innovations have been proposed, tested and, in many cases, implemented in industry. The concepts applied are the following:

- **flue gas energy recovery:**
  - the thermochemical recuperator and regenerator [7,8] (OPTIMELT process), both enhancing recovery efficiency by reformation of natural gas, CH₄ + H₂O → CO + 3 H₂; the net calorific value of CH₄ and the reformed gas mixture typically amount to 10.0 and 12.6 kWh/m³, respectively;
  - pre-heating of oxygen and natural gas (HOTOX process) [9];
  - various batch and cullet pre-heating concepts [5,10–13]

- **combustion and burners:**
  - oxy-fuel combustion; this concept has now reached maturity and been implemented at many sites; it can hardly be referred to as an innovation any longer;
  - top-burner technology [14,15];
  - flameless oxidation technology (FLOX burners) [16]

- **revolutionary melting and fining concepts:**
  - stirred melters [17–19] complemented by a segmented furnace concept (e.g. the RAMAN melter);
  - rotary kilns [17,20,21]; this concept dates back to 1928; it has been supplemented by segmented melting (P10 furnace) [22];
  - cyclone melters (VORTEC) [23];
  - ablation liquefaction concepts [24–26];
  - plasma arc melting [27];
  - submerged combustion technology (SCM) [28–31];
  - batch infiltration technology [32];
  - centrifugal fining [33,34];
  - vacuum fining; based on a 1926 patent by ASAHI Co; recent patent, e.g. [35].

Batch preparation and input concepts extending from conventional pelletizing to sub-mm granulation, thin-layer input [36], core-shell pellets [37,38] and selective batching [39] must not be forgotten. Likewise, the role of sensors and improved furnace control methods play important roles.

The host of different concepts impressively reflects the glass industry’s pursuit of enhanced energy utilization and production efficiency. The above list is by no means comprehensive, and the reader may not see his preferred concept. Some concepts may already be counted among established technologies, and some may even turn out to push (normalized) energy consumption significantly below the magical line of 2800–3000 MJ/t. To date, however, no such technology is at hand. Other concepts resemble the efforts invested during the 1920s-1960s in optimizing the already mature concept of the steam engine railway locomotive. These efforts yielded incremental improvements indeed, but, at the expense of increasingly complex systems – until the development of railway locomotives ultimately moved to electric and Diesel electric traction.

The present paper does not expand on the different concepts listed above. It is devoted, rather, to the basic physics of continuously operating fuel-fired furnaces with supplementary input of electrical energy (boosting). Although the focus is laid on hearth furnaces with combustion and melting spaces brought into thermal contact by a horizontal surface, the derived principles apply to any design in which hot combustion gases transfer heat to a mass flow of batch or melt.

Furnaces are considered for their basic functions as heat exchangers and chemical reactors, which can be achieved by a relatively small set of variables and key indicators. Within the scope of the paper, the issue of mass flow input-output characteristics of the melting space will not be addressed in detail, although the utilization of melting space has a significant impact on furnace performance; see., e.g. [40]. This particular issue will be delegated to a future paper. Parts of the following elaboration have previously been presented by the author [41,42].

First, as an expression of the 1st law of thermodynamics, the general heat balance is elaborated in detail. Then, the principal behavior of a furnace as heat exchanger is analyzed. In this step, process irreversibility (as an expression of the 2nd law of thermodynamics) must be taken into account. This determines the production rate dependence of
furnace performance. In fact, a furnace operates within the “bottle neck” of a narrow optimum between maximum production rate and minimum specific energy consumption. In the next step, the role of the batch is investigated and a number of case studies performed on the industrial scale are presented, revealing the effects of both the energy demand and kinetics of batch-to-melt conversion. Finally, an expression is derived for the ultimate physical limits of energy utilization efficiency by a fuel-fired furnace in general. In general terms, the approach presented is meant to supplement the most successful analyses of furnace behavior via CFD-FEM methods [43–45].

2. Basic physics of glass furnaces

2.1. Heat balance

Heat balance is a time-independent 1st law presentation of furnace performance. It refers to a fixed pull or production rate \( p \), \( p \) in t/h, or to a time average of \( p \). It may be presented in three different ways: (i) as a power balance, \( P \) in kW; (ii) as a balance of energies per 1 t of produced glass, \( H \) in kWh/t, or (iii) as heat fluxes, \( q \) in kW/m\(^2\) per unit heat transfer area \( A \), A in m\(^2\). These quantities are mutually related as \( P = H \cdot p \), \( q = H \cdot r \) with \( r = p/A \). The area-specific pull rate \( r \) is given in units of t/(m\(^2\)-h) or following industrial convention – in t/(m\(^2\)-d). Figure 1 presents the heat balance of a furnace with flue gas heat recovery in terms of power \( P \). The meanings of its individual terms are explained in the legend in Figure 1.

The cornerstone quantities of the heat balance are the power input \( P_{in} \) and the exploited power \( P_{ex} \). Thermodynamically, both quantities are functions of state, i.e. they do not depend on the nature of internal processes such as heat transfer, melting rate, or internal temperature level. The only temperature level involved is the pull temperature \( T_{ex} \), which is the temperature at which the melt leaves the balance space. In industrial practice, \( T_{ex} \) is maintained as a constant within narrow margins. In furnaces equipped with a throat, \( T_{ex} = 1320 \pm 20^\circ C \) is a typical value. When comparing the energy efficiencies of different furnaces, \( T_{ex} \) should be normalized to 1300°C. Both \( P_{in} \) and \( P_{ex} \) can be determined with high accuracy:

\[
P_{in} = P_{in,F} + P_{boost},
\]

where \( P_{in,F} \) and \( P_{boost} \) denote the shares from fuel and electric boosting, respectively. \( P_{in,F} \) depends exclusively on the net calorific value, \( H_{NCV} \) of the fuel, \( H_{NCV} \) in kWh/m\(^3\)ntp and the volume flow \( V_F \) of the fuel, \( V_F \) in m\(^3\)ntp/h. The index ntp denotes “normal conditions” defined as 0°C, 1 atm, which thus differ from the 25°C, 1 bar generally used for thermodynamic standard states. While 1 m\(^3\)ntp translates to 1.106 m\(^3\) in the thermodynamic standard state, energies depend very little on which reference state is chosen. For example, the \( H_{NCV} \) of pure methane amounts to 802.6 and 802.4 kJ/mol for the normal and thermodynamic reference states, respectively. Thus, the balance in Figure 1 may be read as referring to 25°C, 1 bar. The exploited power, \( P_{ex} \) in kW is given by

\[
P_{ex} = H_{ex} \cdot p,
\]
where $H_{ex}$ in kWh/t denotes the exploited heat
\[ H_{ex} = (1 - y_C) \cdot \Delta H_{chem} + \Delta H(T_{ex}), \quad (3) \]
y_C is the fraction of cullet per 1 t of produced glass, $\Delta H_{chem}$ is the standard heat demand of the batch-to-melt conversion (referring to 25°C, 1 bar) and $\Delta H(T_{ex})$ is the heat content of the melt at exit temperature $T_{ex}$ (relative to 25°C). For an accurate determination of $H_{ex}$, for industrial batches of arbitrary composition, please consult item [46]. By virtue of the following relationship:
\[ P_{loss} = P_{wX} + P_{stack} + P_{wh} + P_{wl} = P_{in} - P_{ex}, \quad (4) \]
the cumulative loss $P_{loss}$ is also a function of state. The further analysis of furnace behavior in Sect. 2.3 is based exclusively on the above set of functions of state. The overall efficiency of energy exploitation is as follows:
\[ \eta_{ex} = P_{ex}/P_{in}. \quad (5) \]
Two further balance equations are derived from Figure 1. The power, $P_{flue}$, set free in the furnace amounts to
\[ P_{flue} = P_{in} - P_{re} = P_{fixe} + P_{flue}. \quad (6) \]
At a given production rate, $P_{flue}$ is the power required to sustain the desired glass quality. If, for example, a problem occurs with the heat recovery system such that $P_{re}$ declines, then $P_{in}$ must be adjusted accordingly to sustain a constant $P_{flue}$. The power, $P_{fixe}$, exchanged with the furnace body (comprising batch, melt, and refractory lining) amounts to
\[ P_{fixe} = P_{wX} + P_{ht} = P_{wh} + P_{wl} + P_{ex}. \quad (7) \]
Both Equations (6) and (7) are useful for pinpointing areas of strong or weak performance by a furnace in terms of the following:
- combustion space efficiency $\eta_{flue} = P_{flue}/P_{fixe}$
- heat transfer efficiency $\eta_{ht} = P_{ht}/P_{in}$
- efficiency of heat recovery $\eta_{re} = P_{re}/P_{flue}$
- furnace insulation efficiency $\eta_{fixe} = 1 - (P_{wX} + P_{wl}/P_{in}$

As instructive as these quantities may be, their practical use is limited by the uncertainties by which the individual power terms can be assessed. Due to the complex structure of a furnace, individual wall losses can be determined as rough estimates at best. The remaining quantities require an accurate determination of volume flows of air (or oxygen) and flue gas as well as their actual heat capacities and temperatures. Their assessment is significantly less accurate than the assessment of $P_{in}$ and $P_{ex}$.

The following example refers to air-fuel combustion. $V_{flue}$ and $V_{flue}$ are the volume flows of air and flue gas, respectively, taken under “normal conditions” (ntp). The volume ratios of air to fuel and flue gas to fuel under hypothetical stoichiometric combustion are denoted as $n_{air}$ and $n_{flue}$, respectively. Air excess is quantified by the air excess factor $\lambda$. It is typically set to $\lambda = 1.02$ to 1.08; access of false air has to be taken into account. Thus, the following relationships remain valid:
\[ V_{air} = \lambda \cdot n_{air} \cdot V_{f}', \quad (8a) \]
\[ V_{flue} = (n_{flue} + (\lambda - 1) \cdot n_{air}) \cdot V_{f}', \quad (8b) \]
\[ V_{bg} = (m_{batch} - p) \cdot (v_{ch} / M_{bg}'), \quad (8c) \]
\[ P_{flue} = V_{flue} \cdot c_{flue} \cdot (T_{flue} - T_0), \quad (8d) \]
\[ P_{re} = V_{air} \cdot c_{air} \cdot (T_{air} - T_0), \quad (8e) \]
\[ P_{stack} = V_{flue} \cdot c_{flue} \cdot (T_{flue} - T_0). \quad (8f) \]
Here, $V_{f'}$ is the volume flow of fuel, $m_{batch}$ the mass flow of the batch in t/h, $V_{ch} = 22.4 \text{ m}^3/\text{kmol}$ the molar volume of ideal gasses, and $M_{bg}$ the average mass of batch gases in t/kmol, all taken at ntp. The heat capacities of flue gas and air, $c_{flue}$ and $c_{air}$, respectively, in units kWh/(m$^3$·K) must be calculated from the actual gas compositions and temperatures by using tabulated thermodynamic data and a polynomial expression such as $c = A + B \cdot T + C \cdot T^2 + \ldots$. It is this large number of input data in Equation (8a–8f) and their uncertainties that render the above analysis less accurate than the principal balance Equations (1–5). Nevertheless, valuable conclusions may be drawn: The efficiency of the heat recovery system is constrained to
\[ \eta_{re} = P_{re} / P_{flue} \leq n_{air} / n_{flue} \cdot c_{air} / c_{flue}. \quad (9) \]
The efficiency $\eta_{re}$ is always smaller than its physical limit set by the ratios $n_{air}/n_{flue} \approx 0.9$ and $c_{air}/c_{flue} \approx 0.9$, and hence $\eta_{re} < 0.81$. If the wall losses $P_{wX}$, finite exchange surfaces and finite time operation are taken into account, $\eta_{re}$ settles at around 0.55 to 0.65. Combining Equations (7) and (8a–8f) and solving for $V_{f'}$ yields the following expression for fuel demand:

Through its different terms, this equation summarizes in a pedagogical way the various ways of saving energy by conventional means and based on the 1st law. Clearly, $V_{f'}$ is minimized if

- $\eta_{re}$ is the maximum,
- $V_{bg}$ is low, i.e. batches releasing small volumes of gas are employed,
- the furnace walls are well insulated (although corrosion of the refractories is a factor to be carefully considered),
- $H_{ex}$ is minimized, e.g. by high amounts of cullet or by raw materials with low heat demand in the batch-to-melt conversion,
\[
V_f = \frac{(1 - \eta_{re}) \cdot C_{flue} \cdot V_{bg} + P_{w_m} + P_{w_l} + H_{ext} \cdot \rho}{H_{NCV} = (1 - \eta_{re}) \cdot (n_{flue} \cdot C_{flue} + (\lambda - 1) \cdot n_{air} \cdot C_{air}) \cdot (T_{flue} - T_0)}
\]

where \(T_{ntp} = 273 \text{ K} < T_i\), \(\eta_{melt}\) the average combustion gas temperature in K, \(\rho_{melt}\) the density of the melt in \(\text{t/m}^3\); \(V_{flue}\) is the flue gas volume flow in \(\text{m}^3/\text{h}\) and \(\rho\) the production rate in \(\text{t/h}\). The device sketched in Figure 2 operates among three principal temperature levels:
- the environmental temperature \(T_{0}\),
- the melt exit temperature \(T_{ex}\),
- the adiabatic flame temperature \(T_{ad}\).

The flue gas temperature \(T_{flue}\) by contrast, does not belong to the set of principal temperatures; it settles, rather, as a result of the heat exchange process. The adiabatic flame temperature \(T_{ad}\) is the temperature reached if the entire heat set free in the combustion process were stored in the thermally isolated flue gas. Although this temperature is not reached anywhere in the combustion space, its thermodynamic importance stems from the fact that it marks the temperature of the upper heat reservoir. For exemplary flue gas compositions in vol. %, 72.1 \(\text{N}_2\), 1.4 \(\text{O}_2\), 9.2 \(\text{CO}\), 17.3 \(\text{H}_2\text{O}\) (typical of air-gas combustion) and 4.7 \(\text{N}_2\), 1.3 \(\text{O}_2\), 31.3 \(\text{CO}_2\), 62.7 \(\text{H}_2\text{O}\) (typical of oxy-fuel combustion), \(T_{ad}\) was calculated. In the calculation, species \(\text{N}_2\), \(\text{H}_2\text{O}\), \(\text{CO}_2\), \(\text{O}_2\), \(\text{H}_2\), \(\text{CO}\), \(\text{OH}\), \(\text{O}\), \(\text{H}\), \(\text{NO}\) have to be taken into account. The following equation may be used as a guideline to calculate \(T_{ad}\) in °C in good approximation:

\[
T_{ad} = \theta_1 \cdot (1 + \sqrt{\theta_2 \cdot (H_{NCV} + H_{re}) + \theta_3})
\]

\[
H_{re} = 0.727 + 3.41 \cdot 10^{-3} \cdot T_{re} + 3.31 \cdot 10^{-7} \cdot T_{re}^2
\]

### 2.2. The furnace as heat exchanger

Figure 2 illustrates the function of a glass furnace as a heat exchanger composed of a combustion space with volume \(V_{H}\) and a melting space with volume \(V_{L}\). A hot stream (the hot furnace atmosphere) with a heat capacity flow of \(m'_{H} C_{H}\) passes through the combustion space, and a cold stream (the melt) with a heat capacity flow of \(m'_{L} C_{L}\) passes through the melting space. The direction of flows in Figure 2 is schematic only and does not imply any conclusion concerning their actual direction. Here, \(m'_{H}\) and \(m'_{L}\) are given in units of \(\text{t/h}\) and \(c\) in units of \(\text{kWh}/(\text{t} \cdot \text{K})\). The contact area between the hot and cold stream is \(A\); \(\alpha_{ht}\) in \(\text{kW}/(\text{m}^2 \cdot \text{K})\) is the mean heat transfer coefficient. The nominal dwell times of the hot and cold stream \(t_{H, nom}\) and \(t_{L, nom}\), respectively, are given by

\[
t_{H, nom} = (V_{H}/V_{flue}) \cdot (T_{ntp} / < T_{H}>),
\]

\[
t_{L, nom} = \rho_{melt} V_{L}/\rho,
\]

The diagram of Figure 2 is a schematic representation of a glass furnace as heat exchanger. \(T_{ad}\) adiabatic combustion temperature, \(T_{flue}\) flue gas temperature at exit of the combustion space, \(T_{ex}\) melt temperature at exit of the melting space, \(T_{0}\) environmental temperature; \(V\) volume, \(m'\) mass flow, \(c\) heat capacity; indices \(H\) and \(L\) refer to combustion and melting space, respectively; \(t_{nom}\) = nominal dwell time as calculated from mass flow, mass density and space volume; the arrows do not suggest any particular relative direction of the flows.
The constants $\theta_1$, $\theta_2$, $\theta_3$ are compiled in Table 1 for air-gas, air-oil, and oxy-fuel combustion. In Equation (12b) yielding the heat stored in pre-heated air, $T_{ad}$ must be inserted in units of 1000°C; for oxy-fuel combustion, $H_n = 0$. With $T_{ad} = 1100°C$, $H_{NCV} = 9.9$ kWh/m$^3$, the following typical values of $T_{ad}$ are reached:

- 2440°C for air-gas,
- 2520°C for air-oil,
- 2780°C for oxy-fuel,
- 3200°C for $H_2O_2$ combustion.

The operation of the device in Figure 2 is characterized by three dimensionless key indicators. There are, first, temperature efficiency $\zeta$ defined as

$$\zeta = \frac{T_{ex} - T_0}{T_{ad} - T_0}$$  

(13)

It depends on the three principal temperature levels $T_{ad}$, $T_{ex}$, and $T_0$ alone. For a given combustion mode, it assumes a constant value, i.e. $\zeta = 0.53$, 0.51, and 0.46 for air-gas, air-oil, and oxy-fuel combustion, respectively. In Sect. 2.7, it will be shown that the efficiency of a fuel-fired furnace has the physical upper constraint $\eta_{ex} < \Pi \zeta$. The second indicator is the dimensionless heat capacity flow ratio $z_{HL}$

$$z_{HL} = \frac{\dot{m}_{H} \cdot C_H}{m_{L} \cdot C_L}$$  

(14)

In general terms, $z_{HL}$ compares the power delivered by the hot stream to the power up-take by the cold stream at an infinitely large interface A or at an infinitely high heat transfer rate ($a_{HL} \rightarrow \infty$). An inspection of $z_{HL}$ has direct practical consequences: The optimal heat capacity balance is characterized by $z_{HL} \rightarrow 1$. If a furnace suffers from a heat capacity flow mismatch in which $z_{HL}$ is significantly larger than unity, then the furnace performance cannot be improved effectively by measures directed to the enhancement of the heat transfer conditions (such as use of luminous flames, reduction of a foam layer, etc.). The third key indicator is the dimensionless number of transfer units, NTU:

$$NTU = \frac{a_{HL} \cdot A}{m_{H} \cdot C_H}$$  

(15)

NTU thus characterizes the resistance to heat transfer at contact area A. The above indicators are linked by the relationship $\zeta = f(z_{HL}, NTU)$. Since $\zeta$ is a constant for a given furnace, $z_{HL}$ and NTU cannot be optimized independently; they depend on each other. Figure 3(a,b) shows how $z_{HL}$ depends on the pull rate (normalized to the working point of the furnace; to be explained later), furnace construction, combustion mode and type of fuel. The high efficiency of oxy-fuel furnaces is due primarily to their excellent heat capacity flow match and much less to the high adiabatic flame temperature. Many air-gas end port furnaces reach fair to good balances due to the long dwell times of flue gas in the combustion space. A survey of 17 different air-gas end port furnaces (Figure 4) reveals, however, that this is not always the case. Differences must be attributed to differences in the individual furnace design and to further reasons listed at the end of this paragraph. Side port furnaces typically show larger imbalances due to their short flue gas dwell times. As air-oil combustion generates a smaller flue gas volume than air-gas combustion, a better balance is reached by the former (Figure 3(b); data refer to the same furnace). Operationally, $z_{HL}$ may be manipulated directly by the boost ratio:

$$z_{HL} = \zeta \cdot \frac{P_{in}}{P_{ex}} \cdot \left(1 - \frac{P_{boost}}{P_n}\right)$$  

(16)

Beyond this, $z_{HL}$ depends on the kind of fuel and combustion mode (air-fuel or oxy-fuel); low flue gas volumes reduce $z_{HL}$; the amounts of excess air, false air, and batch gases; and effects like the above.

Indirectly, it is influenced by the combustion space volume, the nominal flue gas dwell time, and the actual combustion space utilization. The success of top burner technology is much more attributable to enhanced combustion space utilization than to local impingements on energy.

Figure 5(a,b) underscores the fact that the primary objective of furnace optimization should be directed towards the heat capacity flow ratio $z_{HL}$. It is only with a well-balanced $z_{HL}$ value that heat transfer enfolds its full effects (Figure 5(a)). Figure 5(b) shows the overall efficiencies of the heat exploitation, $\eta_{ex}$ of 22 different container glass furnaces from various locations worldwide. The survey underscores the dominant role of $z_{HL}$ in furnace performance.

### Table 1

| type of combustion | $\theta_1$ | $\theta_2$ | $\theta_3$ |
|-------------------|-----------|-----------|-----------|
| air-gas           | 1247.3    | 0.1168    | -0.8143   |
| oil-gas           | 1247.3    | 0.1255    | -0.8143   |
| oxy-fuel          | 2062.1    | 0.0253    | -0.1309   |

### 2.3. Process irreversibility and rate dependence

To date, process irreversibility was reflected indirectly only by the dimensionless indicator NTU. The heat balance itself (Figure 1) rests on the 1st law of thermodynamics alone and does not take irreversibility into account. It does not reveal anything about the usefulness of a given amount of heat Q in a process. If an amount of heat Q is used for glass melting, then this amount of heat must be available at the temperature of the process, $T_p$. If it were available at a lower temperature of $T_1 < T_p$ only, then lifting Q to the level of $T_2$ would reduce its useful amount by at least a factor $(1 - T_1/T_2)$. Only the amount $Q(1 - T_1/T_2)$...
would contribute to the power input, $P_{in}$. Entropy may thus be interpreted as a measure of energy degradation. By contrast, electrical power input $P_{boost}$ is always available up to almost 100% thanks to the extremely high Fermi temperatures, $T_e$, of the electrode materials (typically $4 \cdot 10^4$ to $1.2 \cdot 10^5$K); hence, any term $T_1/T_e$ is close to zero.

As a second shortcoming of the 1st law treatment, time is not involved as an explicit parameter. In fact, both energy degradation and finite time constraints constitute fundamental “bottle necks” in any process involving transport phenomena and finite conductivities [47]. The “bottle neck” can be illustrated as follows: Large temperature differences obviously ensure

---

**Figure 3.** a-b. Heat capacity flow ratio $z_{HL}$ as a function of pull $p$ normalized to the respective working points of $p_{86}$ of the furnaces. a: side port, air-gas, 180 m$^2$, Cr-green glass, 15 (above line) and 75% cullet; end port, air-gas, 144 m$^2$, Cr-green glass, 40–50% cullet; oxy-fuel, 158 m$^2$, Cr-green glass, 53% cullet; b: end port, 98 m$^2$, flint glass, 50% cullet, air-gas and oil gas.

**Figure 4.** Heat capacity flow ratio $z_{HL}$ as a function of pull $p$ normalized to the respective working points of $p_{86}$ of the furnaces; survey of 17 different end port air-gas furnaces.
high heat transfer rates and sustain high production rates. On the other hand, an irreversible transfer of heat, $Q$, from temperature, $T_2$ to $T_1$, generates entropy

$$\Delta S = Q \left( \frac{1}{T_2} - \frac{1}{T_1} \right).$$

Thus, minimization of energy degradation calls for small temperature differences, resulting in maximum energy efficiency but minimum

---

**Figure 5.** a-b. a: Efficiency of heat transfer plotted vs. the heat capacity flow ratio $z_{HL}$ for different values of the number of transfer units; averages of values calculated for a simple co- and counter-flow heat exchanger. b: Overall efficiency plotted vs. the heat capacity flow ratio $z_{HL}$, survey of 22 different container glass furnaces; crosses: side port air-gas; circles: end port air-gas; triangles: oxy-fuel.

---

**Figure 6.** a-c. So-called chiller plot (a), power plot (b) and energy plot (c); lines calculated from data of an end port air-gas fired furnace, 152 m$^2$, Cr-green glass, 40% cullet; slope and intercept in the chiller plot and constants $a$ and $b$ calculated by linear regression of the range of normal operation only, ignoring the overpull regime; $P_{in} = $ power input, $P_{ex} = $ exploited power, $\eta_{ex} = $ overall efficiency, $H_{ex} = $ exploited heat of the batch-to melt conversion, $P_{86} = $ working point of the furnace, $H_{in,86} = $ overall heat demand at working point.
production rates. Trade-offs have thus to be found for operating heat exchangers, in general [48,49], and glass furnaces, in particular. Since continuous glass melting is a quasi-stationary process, at least over extended period of time, the time parameter is reflected by the reciprocal production rate, 1/p in h/t, and energy degradation is reflected by the reciprocal efficiency, 1/\eta_{ex} = P_{in}/P_{ex}. The general behavior of a heat exchanger is shown in the "chiller plot" in Figure 6(a). The bold line displays the above-mentioned "bottle neck" behavior, ranging from wastefully low pull across a linear range of normal operation towards an optimum (working point), and then towards wastefully high pull (overpull) with strongly decreasing efficiency. In glass production, this inevitably accompanies quality degradation. Overpull is indeed found in rare cases in industrial practice (see Figure 7; air-gas fired furnace, A = 94 m², Cr-green glass containers; no further information available). Market pressure may have been the reason driving this furnace into overpull. If not analyzed properly, such behavior may even go unnoticed, with only an enhanced occurrence of quality problems observed. The graphs in Figure 6(b) ("power plot") and 6c ("energy plot") are derived directly from Figure 6(a). The data also refer to a real furnace, i.e. an end port air-gas furnace with A = 152 m² and a cullet fraction of 40% producing Cr-green container glass. For practical purposes, this suffices only to evaluate the range of normal operation while ignoring the rare cases of overpull. This yields a straight line P_{in} = a + b·p in the power plot and a hyperbola H_{in} = a/p + b in the energy plot (Figure 6c). Both relations may be familiar to industry personnel; the previous line of reasoning reveals that they are rooted in fundamental principles.

2.4. Data normalization

The procedure outlined in Figure 6(b) is applied to data taken from an end port air-gas furnace, A = 97 m², Cr-green container glass, 70 ± 2% cullet. The result is shown in Figure 8. This time, the principal quantities, P_{in}, P_{ex} and P_{loss}, of the heat balance are shown together. Intercept \( a \) represents the zero pull losses comprising both wall losses and the power required to keep the melt hot; intercept \( b \) is a measure of pull-independent irreversibility. Slope \( b \) is the power increment per unit pull. In an ideal situation, the cumulative losses would feature a horizontal line; hence, \( b \approx H_{ex} \). The slope of the loss line becomes \( (b - H_{ex}) \); this is a measure of pull-dependent irreversibility. Attention is drawn to the excellent statistical precision, i.e. the narrow ±1σ range of the P_{in} line. This is a typical feature of this type of analysis, in spite of the manifold fluctuations in daily factory routine. It is recommended that data collected over extended operation periods of at least ½ to 1 year be retrospectively evaluated, and that the data be sorted according to the (mostly stepwise) changes in cullet, boost ratio, etc. As a result, the pull-dependent performance of a furnace may be characterized most accurately by evaluating the functions of state P_{in} and P_{ex} alone. The fundamental aspects of furnace performance are captured by just three parameters, namely, by \( a, b, \) and \( H_{ex} \).

In the analysis, assessment of the working point of a furnace is important. The working point is the highest pull rate at which – in the operator’s experience – glass of the desired quality is safely produced. In the present paper, a unified procedure is proposed: The working point \( p_{86} \) is defined as the 86% level of all pull rates actually realized, i.e.:

\[
p_{86} = <p> + \sigma_p \tag{17a}
\]

where \(<p>\) denotes the average pull rate and \( \sigma_p \) the standard deviation. If pull rates vary over a very wide range yielding a very large \( \sigma_p \), then \( p_{86} \) is better determined as

\[
p_{86} = p_{\text{MIN}} + 0.86 \cdot (p_{\text{MAX}} - p_{\text{MIN}}). \tag{17b}
\]

If furnaces with different production capacities are to be compared, the pull is normalized to \( p \rightarrow \phi = p/p_{86} \), which is a dimensionless quantity. This procedure has already been used in Figure 3(a,b). In the same way, the power axis may be normalized, to either the heat transferred area \( A \); \( P \rightarrow q = P/A \) in kW/m²; or to the combustion volume \( V_{ch} \); \( P \rightarrow P_{V} = P/V_{ch} \) which is a measurement of power density in kW/m³; or to the power exploited at working point \( p_{86} \); \( P \rightarrow \Phi = P/(H_{ex} p_{86}) \), which is a dimensionless quantity. With the latter normalization step, the effect of batch enthalpy is eliminated. When applying the above normalization procedure to the data

![Figure 7. Chiller plot of an air-gas furnace, 94 m², Cr-green glass, operated in both normal and overpull mode.](image-url)
from an individual furnace with a constant cullet content and boost ratio, the result may appear boring, since the original graph does not change at all. The normalization procedure, however, permits comparison of furnaces of different sizes and production capacities as well as of the effects of different batches, cullet ratios, or boost ratios within the same furnace. Figure 9(a,b) demonstrates this for the example of a side port air-gas furnace, A = 180 m², five different glass colors comprising amber and two distinctly different cullet fractions of 20% (presumably for amber) and 74% (for the others). Figure 9(a) shows the different power demands, $P_{in}$, resulting from the different cullet amounts. As might be expectedly, this is reflected by a significant difference in power demand, $P_{in}$. Note that the slope of the cumulative losses ($b - H_{ex}$) is higher for

Figure 8. Power plot of an end port air-gas furnace, 97 m², Cr-green glass, 70 ± 2% cullet showing power input $P_{in}$, exploited power $P_{ex}$, and cumulative losses $P_{loss}$; different axis show the different options to normalize the pull axis $p$ and the power axis $P$; $H_{ex} =$ exploited heat, $p_{86} =$ working point.

Figure 9. a-b. Power plot (a) and dimensionless power plot (b) of a side port air-gas furnace, 180 m², five different glass colors comprising amber, 20% and 74% cullet.
2.5. Intrinsic heat transfer

In the heat balance (Figure 1; Equation (7)), the power \( P_{ht} \) transferred from the combustion space to the melting space is a critical factor. Due to the large uncertainties at which wall losses, \( P_{wl} \), can be determined, \( P_{ht} \) remains inaccessible by the heat balance. Therefore, an alternative approach is chosen. This rests on a simple (gray = wavelength-independent) radiation model. The features of this model are sketched in Figure 10. In general terms, the incident radiation \( (I) \), absorbed \( (Ab) \) or reflected \( (Rf) \): \( I = Tr + Ab + Rf \). For heat radiation in a furnace, \( Tr \) is approximated as zero and the melt is treated as a surface radiator. This is a fair approximation for amber and Cr-green glass and still a useful one for flint glass. According to Kirchhoff’s law, the absorbed portion is equal to \( Em \) portion emitted as thermal radiation: \( I = Em + Rf \). By Stefan-Boltzmann’s law, \( Em = \varepsilon \sigma T^4 \), where \( 0 < \varepsilon < 1 \) is the emissivity of the radiating volume or surface and \( \sigma \) denotes the universal Stefan-Boltzmann constant \( \sigma = 56.7 \cdot 10^{-12} \text{ kW/(m}^2\cdot\text{K}^4) \). A black body \( (\varepsilon = 1) \) at 1000 K thus emits 56.7 kW/m².

The balance of heat fluxes normalized to the exchange area, \( q = P/A \) in kW/m², constitutes the following equation system:

\[
\begin{align*}
q_1 &= q_{\text{melt}} - T_4^{\text{melt}} + (1 - \varepsilon_{\text{melt}}) \cdot q_2, \quad (18a) \\
q_2 &= q_{\text{gas}} - T_4^{\text{gas}} + (1 - \varepsilon_{\text{gas}}) \cdot q_3, \quad (18b) \\
q_3 &= q_{\text{crown}} - T_4^{\text{crown}} + (1 - \varepsilon_{\text{crown}}) \cdot q_4, \quad (18c) \\
q_4 &= q_{\text{gas}} - T_4^{\text{gas}} + (1 - \varepsilon_{\text{gas}}) \cdot q_1. \quad (18d)
\end{align*}
\]

The emissivities, \( \varepsilon \), of melt, gas, and crown are labeled accordingly. The flux, \( q_{ht} = q_2 - q_1 \), is derived directly from the solution of this equation system, reading

\[
q_{ht} = \frac{a}{b} \left( T_4^{\text{gas}} - T_4^{\text{melt}} \right)^2 + \frac{c}{b} \left( T_4^{\text{crown}} - T_4^{\text{melt}} \right) \cdot q_{\text{ex}} + H_{\text{ex}} \cdot r \quad (19)
\]

where the constants \( a, b, c \) represent combinations of the emissivities involved:

\[
\begin{align*}
a &= \sigma \cdot E_{\text{melt}} \cdot E_{\text{crown}} \cdot (2 - \varepsilon_{\text{gas}} - \varepsilon_{\text{crown}} + \varepsilon_{\text{gas}} \cdot \varepsilon_{\text{crown}}), \quad (20a) \\
b &= \alpha \cdot E_{\text{melt}} \cdot E_{\text{crown}} \cdot (1 - \varepsilon_{\text{gas}}), \quad (20b) \\
c &= 1 - (1 - \varepsilon_{\text{gas}})^2 \cdot (1 - \varepsilon_{\text{melt}}) \cdot (1 - \varepsilon_{\text{crown}}) \quad (20c)
\end{align*}
\]

In Equation (19), Equations (2) and (7) from the heat balance have been invoked, yielding \( q_{ht} = q_{\text{wl}} + H_{\text{ex}} \cdot r \). As crown temperatures are routinely measured in industrial furnace operation, Equation (19) is solved for \( T_{\text{crown}} \), yielding:

\[
T_{\text{crown}} = \left[ \frac{c}{b} \cdot \left( q_{\text{loss}} + H_{\text{ex}} \cdot r \right) - \frac{a}{b} \cdot T_4^{\text{gas}} + \left( 1 + \frac{a}{b} \right) \cdot T_4^{\text{melt}} \right]^{1/4} \quad (21)
\]

Hence, a plot of \( T_{\text{crown}} \) vs. the area specific pull rate \( r \) yields a family of curves with parameters \( H_{\text{ex}} \), \( T_{\text{melt}} \), \( T_{\text{gas}} \) and \( q_{\text{wl}} \) if reasonable values are employed for the

![Figure 10](image-url)
emissivities $\varepsilon$. As a result, parameter studies can be performed to determine what pull rate $r$ is sustained at a given crown temperature $T_{crown}$ under distinct variations in the above parameters. In the following study, $\varepsilon_{crown} = 0.5$ is chosen as a fixed value; $\varepsilon_{gas} = 0.12$, 0.25, and 0.4 are chosen for non-luminous air-gas flames, luminous air-gas flames, and oxy-fuel flames, respectively. The emissivity of the melt surface has been determined experimentally on a 10 kg sample in a lab furnace; the lab furnace is described in [50]. The result of this as yet unpublished experiment is shown in Figure 11. Thus, $\varepsilon_{melt} = 0.2$ is employed for the early stages of batch melting and 0.4 for the later stages in the presence of a foam layer. A clear melt surface approaches a value of $\varepsilon_{melt} > 0.6$.

Figure. 12(a–c) summarize some results of the parametric study; for $q_{net}$, a fixed value of 20 kW/m² is employed. Figure 12(a) shows the strong effect of the gas temperature, even for non-luminous flame ($\varepsilon = 0.12$) over a high-cullet batch ($H_{ex} = 500$ kWh/t). The conditions $\varepsilon_{melt} = 0.4$ and $T_{melt} = 1450°C$ represent the requirement that a sufficiently high fining temperature is reached, even in the presence of residual foam. At a crown temperature of $T_{crown} = 1600°C$, $T_{gas} = 1600°C$ would sustain a pull rate of $r = 2.1$ t/(m²·d) only when $T_{gas} = 1700°C$ would sustain $r = 3.3$ t/(m²·d). Figure 12(b) shows the effect of energy demand $H_{ex}$. In this plot, $T_{gas} = 1650°C$, $\varepsilon_{gas} = 0.25$; the conditions of the melt surface are the same as before. The energy demand of the batch-to-melt conversion has the most significant impact on the sustained pull rate ranging from $r \approx 2.2$ t/(m²·d) for 650 kWh/t (typical of an ECR fiber batch) to 3.3 t/(m²·d) for 405 kWh/t (typical of a batch with very high cullet content). Finally, Figure 12(c) contrasts primary batch melting and fining for an ERC fibre batch (oxy-fuel, hence $\varepsilon_{gas} = 0.4$, $H_{ex} = 650$ kWh/t; $\varepsilon_{melt} = 0.2$ and $T_{melt} = 1250°C$ for the primary batch-to-melt conversion) with a high cullet container glass batch (air-gas with $\varepsilon_{gas} = 0.2$, $H_{ex} = 500$ kWh/t; $\varepsilon_{melt} = 0.2$ and $T_{melt} = 1200°C$ for the primary batch-to-melt conversion). The conditions of the melt surface in the fining area are the same as before. In both cases, the requirement to reach a sufficiently high fining temperature imposes a tighter constraint on the achievable pull rate than batch melting. This result requires careful discussion. The findings in Figure 12(c) apply if heat transfer is the rate-controlling step during batch melting, i.e. if the chemical turnover rate instantaneously follows heat transfer. In batches with sluggish chemical turnover rates, batch melting may become the critical step constraining the pull rate.

2.6. Role of the batch

The last finding draws attention to the batch. Clearly, furnace performance depends on the exploited heat $H_{ex}$, hence, on the enthalpic condition of the batch. Figure. 9(a,b) have already shown that, besides the enthalpic effect, the kinetics of the batch-to-melt conversion also have a significant impact. Figure 12(c) shows that batch kinetics are insignificant only if the chemical conversion rate can follow the heat transfer rate. This may not be the case at all during the early stages of batch melting. Therefore, optimization of the batch with respect to both low energy demand and high chemical conversion rates offers a huge potential for furnace performance optimization.

Standard chemical heat demand $\Delta H_{chem}$ for batches yielding an identical glass composition (insulation glass wool) but using different boron oxide carrier raw materials is compiled in Figure 13. The enthalpy differences between the raw material options are significant. The time demand of the batch-to-melt conversion was determined for the 10 kg lab furnace mentioned earlier [49]. A batch layer of 5 cm thickness charged on a preheated cullet melt maintained at 1200°C was exposed to heat radiation of 1400°C. The observed conversion times were $770 \pm 10$, $1510 \pm 55$, $1300 \pm 10$, and $1445 \pm 15$ s for the pentaborate, tincal, colemanite, and ulexite batches, respectively. Both the enthalpic and kinetic effects played a role; in the study discussed, they exerted an antagonistic effect in some cases: Note that the pentaborate batch with its relatively high energy demand displayed a conversion time 50% shorter than the
colemanite batch with its low energy demand. It is thus difficult to predict how a raw material replacement would impact the performance of a furnace. In a later factory campaign (production rate \( r < 2 \) t/m\(^2\)·d; furnace data not specified here), a replacement of pentaborate with ulexite could indeed be implemented successfully. At the furnace’s low production rate, the slow conversion rate of the ulexite batch did not reveal any negative effects. Even a shorter foam layer had observed. The enhanced carry-over of the ulexite batch due to the very small grain sizes of the available raw material was a nasty effect.

In the following paragraphs, case studies are presented in which the effects of raw material substitution at a constant glass composition were studied on an industrial scale. The campaigns were performed according to the following scenario: A production period with a constant production rate was depicted: A 7 d period with the original batch was followed by a 7 d period with the alternative batch, and then by a 7 d period with the original batch again.

**CASE 1:** In a flint glass production experiment, dolomite was replaced by half-burnt dolomite. The heats of formation from the elements of these raw materials amounted to 2342 and 1809 kJ/mol, respectively. At a 4 wt. % MgO level in the glass formula and upon use of 40% cullet, this yields an intrinsic energy advantage of \( \Delta H_{ex} = 18.6 \) kWh/t. The test campaign was performed in parallel in two end-port air-gas furnaces of very similar construction with \( A = 40 \) and 40.5 m\(^2\) for furnaces I and II, respectively. Furnace II
had smaller regenerator chambers than furnace I. Figure 14(a–c) summarize the results. The two furnaces differed considerably in their performance. Pull-independent irreversibility (intercept a) was higher in furnace I, while pull-dependent irreversibility (the slope of the loss line, b = H_{ex}) was higher in furnace II. The heat capacity flow ratio z_{HL} was 1.16 vs. 1.21; efficiencies, \eta_{ex}, at the working point (4.5 vs. 5.2 t/h) were 45% and 44%, closely consistent with the results in Figure 5(b). The raw material substitution yielded the expected decrease of 40 kWh/t in furnace I, while a much smaller effect was found in furnace II. From this campaign, we may learn that the effect of low enthalpy batches should be tested only in furnaces with a small loss slope b = H_{ex} and that furnaces with a large loss slope react quite insensitively to such changes.

CASE 2: At another production site, limestone was replaced by lime, Ca(OH)$_2$. The intrinsic energy was reduced by \Delta H_{ex} = 33 kWh/t. A campaign in a small end port-fired furnace, A = 35 m$^2$, producing amber bottles (Figure 15) yielded a decrease in the overall energy demand of approx. 38 kWh/t from H_{in} = 1102 ± 11 to 1064 ± 11 kWh/t. When taking into account the efficiency \eta_{ex} of this furnace, a larger decrease should be expected eventually. However, during a 7 d campaign, a cautious operator cannot explore the limits.

CASE 3: Two lead crystal glass batches were investigated in a small all-electric furnace producing stemware (cold-top, A = 8 m$^2$). Although lead has been largely eliminated from today’s glass making, this case is reported for its insight into the role of batch conversion kinetics. The batches were identical with respect to both chemical composition and mineral phase content. Consequently, at constant T_{ex}, there were no differences in H_{ex}. However, one batch contained conventional raw materials, while the other consisted of sub-mm composite granules. During the campaign (see Figure 16(a)), the production rate could be increased by 18% from 340 to 400 kg/h without any loss of quality. Along with the production rate, the overall energy demand decreased by 8% from 863 to 797 kWh/t. This strong effect was observed in spite of the fact that the pull temperature T_{ex} increased from 1265°C to 1285°C (Figure 16(b)); at a constant T_{ex}, the positive effect would have been even higher. In another factory, the same pair of batches was tested again. Here, the operator did not respond to a request to increase the pull rate cautiously. As expected, no difference between the two batches was observed in that case.

As illustrated by the above cases, the enthalpy effects of a batch usually translate to the industrial scale in a straight-forward manner. In simple batch-free time lab experiments, cases where a low-enthalpy batch responds with a retarded chemical conversion rate are readily identified (as in the pentaborate vs. colemanite case). The actual outcome in industrial application depends on the applied production rate as well as on the furnace’s rate-dependent irreversibility as reflected by the loss slope, b = H_{ex}. A large irreversibility diminishes or even annihilates the otherwise expected positive effects. Fast conversion batches extent a strong

![Figure 14.](image-url)
positive effect on the achievable production rate; by virtue of this effect alone, the overall energy demand is lowered. At a maintained pull, little or no effect is observed in these cases.

Finally, a further way of intrinsically optimizing the melting process is addressed. This approach also involves the glass composition. It is the design of low-liquidus glasses. The liquidus temperature $T_{\text{liq}}$ of a glass melt is the temperature above which no crystals may precipitate from the melt. Along the time direction of the batch-to-melt turnover, $T_{\text{liq}}$ marks the temperature level below which sand dissolution cannot be completed. Thus, a high $T_{\text{liq}}$ value narrows the temperature window, and hence, the time window for sand dissolution. As any dissolving sand grain
leaves a tiny bubble cluster behind in the melt, a lag of sand dissolution also narrows the time window for fining. Thus, a low-liquidus glass formula would reduce the dwell time of the melt required to reach the desired glass quality. The effect on furnace performance would be similar to the effect of using a fast conversion rate batch. It is true that container glass producers maintain stringent control over the viscosity-temperature curve, especially, of the gob temperature, $T_{3.0}$ (i.e. the temperature at the $10^{3.0} = 1000$ dPa·s level). A survey was performed comprising 21 different container glass manufacturers. All of them produced container glass successfully. The gob temperature $T_{3.0}$ varied within a narrow margin of just a few degrees centigrade. By contrast, the liquidus temperatures varied between 980°C and 1085°C. The values of $T_{liq}$ did not correlate with the amount of soda ash in the batch. Among the lot, there were soda ash-poor batches with low $T_{liq}$ and very soda ash-rich batches with high $T_{liq}$. The allocation of $T_{liq}$ – if there was any intentional allocation at all – did not seem to respond to a need stemming from the raw material situation nor to serve any other particular purpose. In view of the fact that the above variation width of $T_{liq}$ corresponds to a variation in sand dissolution time of approx. ±20%, it is strongly recommended that $T_{liq}$ be taken into serious consideration for optimization of conventional container glass formulae and batches. The perspective of minimizing the use of soda ash should be a strong economic motivation. The compositional dependence of $T_{liq}$ is, however, not easily determined, either theoretically or experimentally. Simple oxide increment calculation schemes for $T_{liq}$ are of little use. The calculation of $T_{liq}$ instead requires the use of a CALPHAD-type calculation program [51] in combination with a careful experimental determination of $T_{liq}$. This issue has been addressed previously for float glasses [52].

### 2.7. The physical limits of fuel-fired furnaces

In view of the CO$_2$ emissions by combustion of fossil fuels, all-electric melting has been proposed as the future technology of glass melting (provided that the electricity is obtained from renewable resources). On the other hand, the use of “green” fuel from renewable resources would yield the same positive environmental effects. In this context, the question arises as to what limits of energy exploitation efficiency a fuel-fired furnace, supported by electrical boosting, can be driven from a theoretical point of view. For this purpose, a process similar to the Curzon-Ahlborn (C.A.) cycle [53] is discussed (see Figure 17). It is true that the C.A. cycle focuses on the generation of power using a heat exchange process. But if we assume that entropy generation in the melting space is small compared to entropy generation in the combustion space and during heat transfer, this scenario can be applied to glass melting in an approximate way. The power, $P_{ht}$, transferred to the cold stream must be sustained by the limited upper reservoir of the combustion space. This is expressed by the relationship:

![Figure 17. Sketch of the finite time heat transfer in analogy to a Curzon-Ahlborn cycle [53]; indices have the same meaning as in Figure 1; $T_{ad}$ = adiabatic flame temperature.](image-url)
where $K$ is a transfer coefficient in kW/(m$^2$·K). The exploited power can then be given by

$$P_{\text{ex}} = \frac{P_{\text{ht}}}{C_1 T_{\text{ht}}}$$

(23)

At very low pull rates, $T_{\text{ht}} \to T_{\text{ad}}$, while at very high rates, $T_{\text{ht}} \to T_{\text{ex}}$. In both cases, $P_{\text{ex}} \to 0$. This is simply an alternative presentation of glass melting’s “bottle neck” problem addressed in Sect. 2.3. Optimal conditions are found by applying the condition $\partial P_{\text{ex}}/\partial T_{\text{ht}} = 0$ to yield

$$T_{\text{ht}} - T_0 = \sqrt{(T_{\text{ad}} - T_0) \cdot (T_{\text{ex}} - T_0)}$$

(24)

and

$$\eta_{\text{ex,max}} = \sqrt{T_{\text{ex}} - T_0} / \sqrt{T_{\text{ad}} - T_0} = \sqrt{\zeta}$$

(25)

Figure 18(a) presents a cumulative “chiller plot” analysis of all furnace data available to the author. For the sake of simplicity, no distinctions are made with respect to furnace construction, combustion mode, or cullet content. A regression analysis of $1/\eta_{\text{ex}}$ vs. $1/r$ is performed; the $\pm 1\sigma$ and $\pm 3\sigma$ error margins are shown by thin straight lines. The intercept settles at a value very close to $\sqrt{\pi \zeta}$; the arrows to the left mark the tiny differences between the actual $\sqrt{\pi \zeta}$ values of air-fuel and oxy-fuel combustion. As a result, $\eta_{\text{ex,max}} = \sqrt{\pi \zeta}$ appears as the physical limit of fuel-fired furnaces, which is in agreement with the C.A.-type analysis. The regression analysis also yields a slope, $r_0 \approx 3$ t/(m$^2$·d), as a rate constant typical of all these furnaces. When adopting the limit, $\eta_{\text{limit}} = \eta_{\text{ex,max}} = \sqrt{\pi \zeta}$, and applying a $\eta_{\text{ex}}$ alternative type of rate dependence (Figure 18(b)), the same value for $r_0$ is found.

An admittedly somewhat speculative physical interpretation is given for the rate constant $r_0$. The rate constant $r_0$ is interpreted by the local constraints of the batch reactions as

$$r_0 = \beta \cdot \rho$$

(26)

where $\rho \approx 2.5$ t/m$^3$ is a typical density of the batch material and $\beta = D/L$ is a mass transfer coefficient in m/s. In the early stages of batch melting (1200°C), $D = D(\text{Na}^+) \approx 2 \cdot 10^{-9}$ m$^2$/s and $L \approx 150$ μm as a typical median radius of glass grade sand grains. This yields $\beta \approx 1.33 \cdot 10^{-7}$, m/s = 1.15 m/d and $r_0 = 2.9$ t/(m$^2$·d). The order of magnitude of this rough estimate is in good agreement with the value for $r_0$ found by data regression. This may be viewed as a hint that it is the same type of local mechanism as in the batch, rather than a macroscopic effect, which constitutes a uniform value for $r_0$ for so many different cases.

### 3. Conclusion

The present paper is devoted to an analysis of the basic physics of continuously operating fuel-fired furnaces with supplementary boosting. This type of furnace still dominates the glass industry and – in view of the long lifetime of modern furnaces – will...
do so for quite some time. It is the purpose of this paper to equip glass manufacturers with a tool to monitor, compare, and optimize furnace performance based on a retrospective analysis of operation data which are collected in any cases during the daily production routine, and to provide additional guidelines for furnace construction. It is meant to complement the widespread and successful use of CFD-FEM analysis, and to compensate for some of its shortcomings. The main results of the paper are summarized below:

The heat balance is a time-independent 1st law presentation of furnace performance. The usefulness of this balance depends on its accuracy. In this respect, two terms in the balance stand out from the rest. These are the power input, \( P_{in} \), and the exploited power, \( P_{ex} \). Both terms are functions of state, and hence do not depend on any internal process. They can be determined with high accuracy. By virtue of their difference, the cumulative losses \( P_{loss} = P_{in} - P_{ex} \) are in the same category. In terms of precise quantification, the usefulness of these other terms is limited by the uncertainties that beset their determination and thus make their detailed evaluation difficult. Nevertheless, they give useful guidelines for furnace optimization, most of which are well known. The demand of fuel input is minimized if

- the efficiency of the flue gas heat exchanger is maximized;
- the heat contained in the flue gas leaving the heat exchanger is exploited, e.g. for pre-heating the batch;
- the flue gas temperature is kept low at the exit of the combustion space, which is basically a matter of furnace construction and burner technology;
- the furnace walls are well insulated, with careful consideration of the impact on refractory corrosion, especially in the melting space;
- the enthalpy of the batch-to-melt conversion is low;
- the amount of gases introduced by the batch is low;
- the air (or oxygen) excess is kept low and influxes of false air are avoided.

The heat balance does not, however, tell how these measures influence each other. The results of individual measures cannot simply be added up. Rather, the stringent optimization of a single aspect may drive other terms into an unfavorable situation.

The function of a furnace as heat exchanger is exploited, e.g. for pre-heating the batch, temperature, \( T_{ex} \), and the environmental temperature, \( T_0 \), among which the furnace operates, second, the heat capacity flow ratio \( z_{HL} \) of the hot and cold streams, and, third, the number of transfer units, NTU, reflecting the heat transfer from the hot to cold stream. Among these quantities, \( z_{HL} \) has an extremely large impact on furnace performance. Only in well-balanced furnaces with \( z_{HL} \rightarrow 1 \), good heat transfer conditions unfold their full effect. Hence, only well-balanced furnaces achieve high efficiencies of energy exploitation. The \( z_{HL} \) ratio depends on

- the combustion mode (air or oxygen) and the kind of fuel used (low flue gas volumes reduce \( z_{HL} \));
- the boost ratio (high boost ratios reduce the flue gas volume);
- the presence of excess gases from either the air factor \( \lambda \) or the volume of batch gases or else from the influx of false air (low volumes of excess gases reduce \( z_{HL} \));
- the dwell time of flue gases in the combustion space; this is an indirect effect based on the mutual linkage of the quantities as \( \zeta = f(z_{HL}, \text{NTU}) \); the dwell time is influenced by the furnace construction and burner technology.

Temperature differences between hot and cold streams are a critical issue. Large temperature differences yield high heat transfer rates and high production rates but drive the process towards high entropy generation. By contrast, small temperature differences minimize entropy generation and the efficiency of energy exploitation, but result in low production rates. This constitutes a "bottle neck" for optimization of the glass melting process. A 2nd law treatment based exclusively on the functions of states \( P_{in} \) and \( P_{ex} \) yields a surprisingly simple route towards furnace performance analysis in terms of three performance parameters; \( a \) (the cumulative losses at zero pull), \( b \) (the power demand per unit pull \( p \)), and the exploited heat \( H_{ex} \) (reflecting the energy demands of the batch-to-melt conversion). Parameter \( a \) presents the pull-independent irreversibility of the process, while \( (b - H_{ex}) \) stands for the pull-dependent irreversibility. \( H_{ex} \) is determined from the batch and glass composition, and \( a \) and \( b \) are determined with fairly high precision from a retrospective analysis of furnace data, preferably collected over periods > \( \frac{1}{2} \) year. Based on these findings, a furnace analysis procedure is proposed which permits direct comparison of furnaces of different sizes and production capacities to evaluate the effects of different batches, and to distinguish between enthalpic and kinetic effects of the batch-to-melt conversion.

A simple radiation model is proposed to assess the effects of gas temperature, intrinsic heat demand and gas and melt surface emissivity on the production
rate. The role of batch with respect to both energy demand and the chemical conversion rate is demonstrated by the data obtained through four selected industry campaigns. Batch optimization appears to be a powerful approach to enhancing furnace operation.

Finally, the physical limits on the efficiency of a fuel-fired furnace in general are derived. The efficiency is given by the square root of the temperature efficiency, $\eta_T$.

**Disclosure statement**

No potential conflict of interest was reported by the author.

**References**

[1] Beerkens RCG, Limpt HV. Energy efficiency benchmarking of glass furnaces. Ceram Eng Sci Proc. 2002;23:93–106.

[2] Beerkens RCG, Limpt HV, Jakobs G. Energy efficiency benchmarking of glass furnaces. Glass Sci Technol. 2004;77:47–57.

[3] Beerkens: RCG. New concepts for energy efficient & emission friendly melting of glass. Ceram Trans. 2012;231:5–24.

[4] Nebel R. Potential development for saving energy and reducing pollutant emission. Proc. Int. NOVEM Workshop on Energy Efficiency; Amsterdam; 2000.

[5] Barklage-Hilgefort H-J. Three years of industrial experience with a cross fired regenerative furnace with batch pre-heater (in German). Oral presentation, Technical Committee II, The German Society of Glass Technology DGG, Offenbach 1998. Glastech. Ber. Glass Sci. Technol. 72 (1999), No. 2, N19.

[6] Hibscher CW, Davies PRH, Davies MP, et al. A designer’s insight into all-electric melting. Proc. 65th Conf. on Glass Problems; Columbus, Ohio; 2004.

[7] Kobayashi H. Thermochemical regenerative heat recovery process. US Patent 6 113 874. 2000.

[8] de Diego J, Kobayashi H, Laux S, et al. Optimel set to reduce emissions at Libbey’s oxy-fuel fired furnace. Glass Int. 2019;40:75–78.

[9] Joumani Y, Jarry L, Simon J-F, et al. Results of a technology recovering waste heat to preheat oxygen and natural gas for ox-y-fuel furnaces. GMIC glass-technology and waste heat management in the glass industry. GMIC, Columbus, Ohio; 2010.

[10] Lindig: M. Efficiency consideration on the principle of batch pre-heating (in German). In: HVG colloquium. Batch: Preparation, melting technology, glass quality. The German society of glass technology, Offenbach. 2011. 11 (1): 38.

[11] Beerkens: RCG. Energy saving options for glass furnaces & recovery of heat from their flue gases - and experiences with batch & cullet pre-heaters applied in the glass industry. Proc. 69th Conference on Glass Problems; Columbus, Ohio; 2008.

[12] Lubitz G, Beutin E, Leimkühler J. Oxy-fuel fired furnace in combination with batch and cullet preheating. Proc. Int. NOVEM Workshop on Energy Efficiency; Amsterdam; 2000 p. 69–78.

[13] Kobayashi S, Evenson E, Miclo E. Development of an advanced batch/cullet preheater for oxy-fuel fired glass furnaces. Glass Trend workshop on Energy Efficiency and Environmental Aspects of Industrial Glass Melting; Versailles, France; 2007.

[14] Lauwers EJ. Roof-mounted auxiliary oxygen-fired burner in glass melting furnace. US Patent S 139 558. 1992.

[15] LeBlanc J, Marshall R, Prusia G, et al. The BOC convective glass melting system. Ceram Eng Sci Proc. 2002;23:107–188.

[16] Scherello A, Gröner K, Konold U, et al. Application of flameless oxidation in glass melting furnaces – glass-FLOX®. Proc. 7th Europ. Conf. on Industrial Furnaces and Boilers; Porto, Portugal; 2006.

[17] Barton: JL. Innovation in glass melting. Glass Technol. 1993;34:170–177.

[18] Sakamoto O. Innovative energy saving glass melting technology. Res Reports Asahi Glass Co Ltd. 2009;59:55–60.

[19] Richards RS. Rapid glass melting and refining system. Proc. Advances in the Fusion of Glass; Alfred, New York; 1998 p. 1–11.

[20] Devilliers RW, Vaerwyck: FE. Le Four a Bassin. Brussels: Les études des composés siliceux, 1935.

[21] Kobayashi 270H, Leblanc JR. Oxy-fuel fired rapid glass melting processes. Proc. Advances in the Fusion of Glass; New Orleans; 1992.

[22] Demarest HM, Mataes JM, Kunkle GE. Method of calcining and liquefying glass batch. US Patent 4 539 030. 1985. ibid: Method of Melting Raw Materials for Glass or the Like with Staged Combustion and Preheating. US Patent 4 634 461 (1987).

[23] Bender DJ, Hnat JG, Litka AF, et al. Pilot-scale testing and preliminary commercial system design of a gas-fired advanced glass melting furnace. Ceram Eng Sci Proc. 1990;11:102–124.

[24] Kunkle GE, Mataes SM. Glass batch liquefaction. US Patent 4 381 934. 1983. ibid: Methods for Ablating Liquefaction of Raw Materials. US Patent 4 564 379 (1986); ibid: Apparatus and Method for Ablating Liquefaction of Materials. US Patent 4 654 068 (1987).

[25] Demarest HM Jr.: Two stage batch liquefaction process and apparatus. US Patent 4 519 814. 1985.

[26] Henry G Process for melting glass and the like. US patent 2 634 555. 1953.

[27] Gunterman R, Weinstein MA. High intensity plasma glass melter project. DOE (US Department of Energy) Report DE-FC36-03GO13093. 2006.

[28] Rue D. Submerged combustion melting. Ceram Bull. 2004;10:18–20.

[29] Rue D, Kunc W, Aronchi G Operation of a pilot-scale submerged combustion melter. Proc. 68th Conference on Glass Problems; Columbus, Ohio; 2007 p. 125–135.

[30] Rue D, Brown: JT. Submerged combustion melting of glass. Int J Appl Glass Sci. 2011;2:262–274.

[31] Purnode B Submerged combustion glass manufacturing system and method. US patent 9 481 592. 2009.

[32] Yao Y, Watanabe T, Yano T, et al. An innovative energy-saving in-flight melting technology and its
application in glass production. Sci Technol Adv Mater. 2008;9:025013.

[33] Nemec L, Tonarova V. Glass melting and its innovation potentials: bubble removal under the effect of the centrifugal force. Ceram Silik. 2008;52:225–239.

[34] Tonarova V, Nemec L, Jebava M. Bubble removal from glass melts in a rotating cylinder. Eur J Glass Sci Technol. 2010;A51:165–171.

[35] Kawagushi T, Obayashi K, Okada M, et al. Vacuum degassing method for molten glass flow. US patent 6 332 339 B1. 2001.

[36] Wiltzsch S, Hessenkemper H. First experimental and numerical simulation results on a segmented thin-layer melter. The German Glass Society (DGG), Annual Meeting: Saarbrücken; 2011.

[37] Leese SM, Stuart JA. Patent WO2012/101429 A2.

[38] Conradt R, Prange A, Niessen V, Ross C. Patent WO2016/110314 A1.

[39] Carty WM, Kim U, Sinton CW. Selective batching for improved commercial glass melting. Ceram Bull. 2004;10:28–32.

[40] Nemec L, Jebava M, Dyrckova P. Glass melting phenomena, their ordering and melting space utilization. Ceram Silik. 2013;57:275–284.

[41] Conradt R. Thermal versus chemical constraints for the efficiency of industrial glass melting furnaces. Ceram Trans. 2012;231:25–36.

[42] Conradt R. The relation between furnace efficiency and the physics and chemistry of the melting process. Ceram Eng Sci Proc. 2018;39:221–231.

[43] Muyssenbergh P, Simonis F, Roosmalen R. Verification of 3D mathematical simulation with measured temperature profiles during furnace operation. Glastech Ber. 1995;68C2:55–62.